OPERATIONAL RISK ASSESSMENT OF ROUTING FLARE GAS TO BOILER FOR COGENERATION

A Thesis

by

SANKHADEEP SARKAR

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Chair of Committee,	Mahmoud El-Halwagi
Co-Chair of Committee,	Chad Mashuga
Committee Members,	Jerome Schubert
Head of Department,	M. Nazmul Karim

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ABSTRACT

Flaring is a controlled combustion process in which unwanted or excess hydrocarbon gases are released to flare stack for disposal. Flaring has a significant impact on environment, energy and economy. Flare gas integration to cogeneration plant is an alternative to mitigate flaring, benefiting from utilizing waste flare gas as a supplemental fuel to boilers and or gas turbines. Earlier studies have shown the energy and economic sustainability through integration. However, the impact of flare gas quality on cogeneration plants are yet to be identified.

This paper studies the effect of flare gas composition and temperature from an ethylene plant to an existing boiler during abnormal flaring. The study proposes a unique framework which identifies the process hazards associated with variation in fuel conditions through process simulation and sensitivity analysis. Then, a systematic approach is used to evaluate the critical operational event occurrences and their impacts through scenario development and quantitative risk assessment, comparing a base case natural gas fuel with a variable flare gas fuel.

An important outcome from this study is the identification of critical fuel stream parameters affecting the fired boiler operation through process simulation. Flare stream temperature and presence of higher molecular weight hydrocarbons in flare streams showed minimal effect on boiler condition. However, hydrogen content and rich fuel-air ratio in the boiler can affect the boiler operating conditions. Increase in the hydrogen content in flare to fuel system can increase the risk contour of cogeneration plant, affecting

the boiler gas temperature, combustion mixture and flame stability inside the firebox. Quantitative risk analysis through Bayesian Network showed a significant risk escalation. With 12 hours of flare gas frequency per year, there is a substantial rise in the probability of occurrence of boiler gas temperature exceeding design limit and rich fuel mixture in the firebox due to medium and high hydrogen content gas in flare. The influence of these events on flame impingement and tube rupture incidents are noteworthy for high hydrogen content gas. The study also observed reduction in operational time as the hydrogen content in flare gas is increased from low to high.

Finally, to operate fire tube steam boiler with flare gas containing higher amount of hydrogen, the existing cogeneration system needs to update its preventive safeguards to reduce the probability of loss control event.

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CONTRIBUTORS AND FUNDING SOURCES

Contributors

This work was supervised by a thesis committee consisting of advisor Professor Mahmoud El-Halwagi of the Department of Chemical Engineering, Professor Chad Mashuga of the Department of Chemical Engineering and Professor Jerome Schubert of the Department of Petroleum Engineering. The initial idea for the thesis was provided by Professor Mahmoud El-Halwagi of the Department of Chemical Engineering.

All the work conducted for the thesis was completed by the student independently

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NOMENCLATURE

AFR	Air to Fuel Ratio
AG	Associated Gas
APG	Associated Petroleum Gas
ARAMIS	Accidental Risk Assessment Methodology for Industries
BMS	Burner Management System
BN	Bayesian Network
BPD	Barrels Per Day
BSEE	Bureau of Safety and Environmental Enforcement
BTU	British Thermal Unit
CBA	Cost Benefit Analysis
CCS	Carbon Capture and Sequestration
CDM	United Nations Clean Development Mechanism
CHP	Combined Heat and Power
CO_2	Carbon Dioxide
CO	Carbon Monoxide
CPT	Conditional Probability Table
EOS	Equation of State
EPA	United States Environmental Protection Agency
FGRS	Flare Gas Recovery Systems
FMEA	Failure Mode and Effect Analysis

FT	Fisher-Tropsch
FTA	Fault Tree Analysis
GGFR	Global Gas Flaring Reduction
GHG	Greenhouse Gas
GMI	Global Methane Initiative
GTL	Gas-To-Liquid
H_2S	Hydrogen Sulfide
HAZOP	Hazard and Operability Study
HHV	Higher Heating Value
HRSG	Heat Recovery Steam Generator
IOGP	International Association of Oil and Gas Producers
IRR	Internal Rate of Return
kWh	Kilowatt hour
LHV	Lower Heating Value
LNG	Liquified Natural Gas
LPG	Liquified Petroleum Gas
MMSCFD	Million Standard Cubic Feet Per Day
MW	Megawatt
PIF	Performance Influencing Factors
PLC	Programmable Logic Controllers
PGPP	Power Generation Partner Program
PR-BM	Peng-Robinson-Boston Mathias

SIS	Safety Instrumented Systems
RIF	Risk Influencing Factors
TE	Top Event
TMD	Thermal Membrane Distillation
TPD	Tons Per Day
TRIT	Turbine Rotor Inlet Temperature
WI	Wobbe Index

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1. INTRODUCTION

Flaring is a controlled combustion process in which unwanted or excess hydrocarbon gas during normal operation (e.g. off specification product streams) or abnormal operation (e.g. plant emergency shutdowns, system over pressurization) are sent to flare stack for complete combustion. Figure 1 shows an overview of flare system in a hydrocarbon processing plant. Generally, flare systems are classified based on the height of the flare tip (elevated or grounded) and method of mixing hydrocarbons at the flare tip (steam, air, pressure assisted or non-assisted) (Leslie B Evans, 2000).

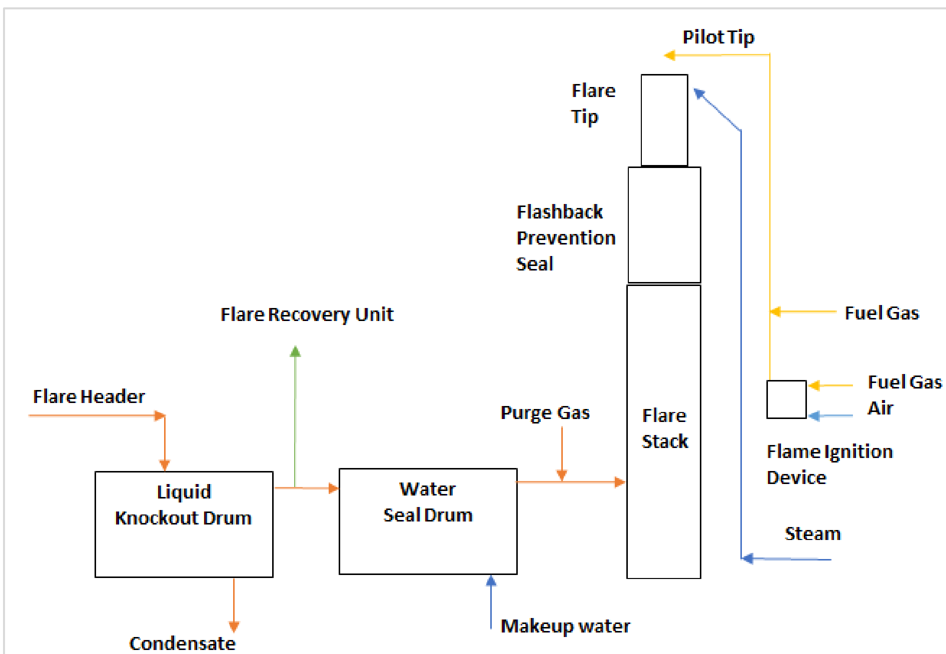


Figure 1 Standard Flare System (Emam, 2015)

Flare systems are prevalent in all the industries encompassing chemical plants, petrochemicals, refineries, onshore and offshore oil and gas platforms. The major challenges of flaring faced by these industries are its environmental and economic impacts. Gas flaring is a significant contributor to greenhouse emissions. During high volume of flaring, unburnt carbons along with greenhouse gases are released to the atmosphere. Annually around 145 billion cubic meters of gas is flared in the world, which shows nearly 300 million metric tons of CO_2 is released to the atmosphere annually (World Bank, 2018). World CO_2 emissions from flaring encompasses more than 50% of the annual Certified Emissions Reductions (624 Metric Tons CO_2) under Clean Development Mechanism (CDM is one of the systems developed under Kyoto Protocol to United Nations Framework on Climate Change (Emam, 2015). Figure 2 shows top ten flaring countries in 2013 and 2017 (The World Bank, 2018). Moreover, financial loss due to flaring are estimated to be around 10-15 billion dollars annually (Farina, 2010).

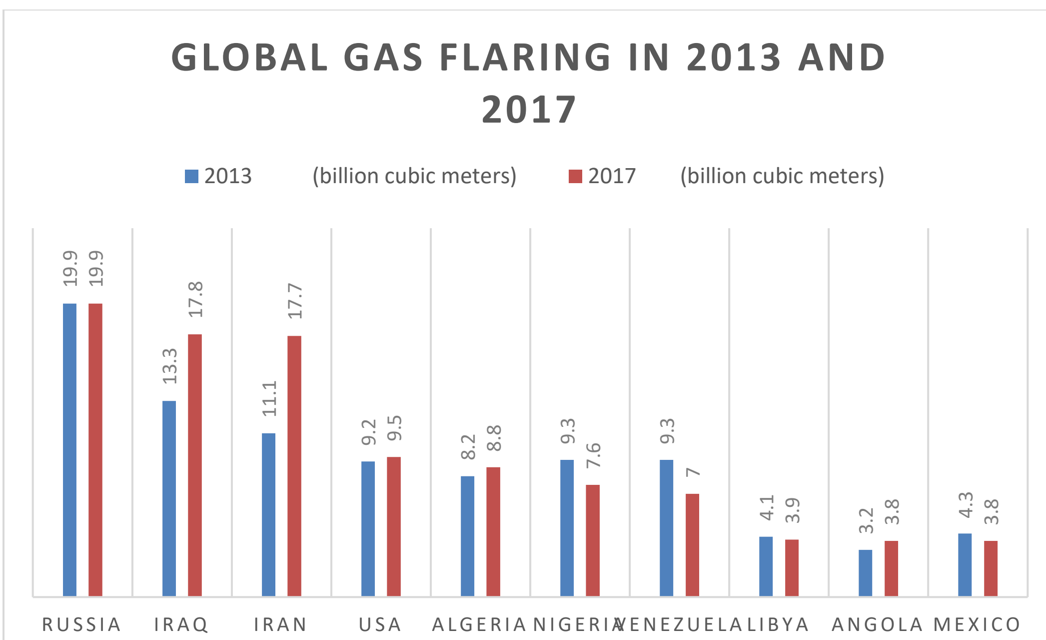


Figure 2 Top Ten Flaring Countries (World Bank, 2018)

Therefore, optimizing fuel consumption, saving energy and reducing gas emissions are the primary goal of every oil and gas industry. One of the approaches of utilizing the energy rich flare gas is through flare gas recovery method.

In cogeneration system, 90% of the operational cost is due to the fuel and maintenance cost is 30% of the fuel cost (Monzure-Khoda Kazi F. M., 2015). Kamrava et al. demonstrated that 90% of CO_2 can be reduced if the flare gas is reutilized (offsetting fresh fuel consumption) in cogeneration system (Serveh Kamrava K. J., 2015). So, using high energy content flare gas as a fuel reduces operational cost of cogeneration systems and CO_2 emissions.

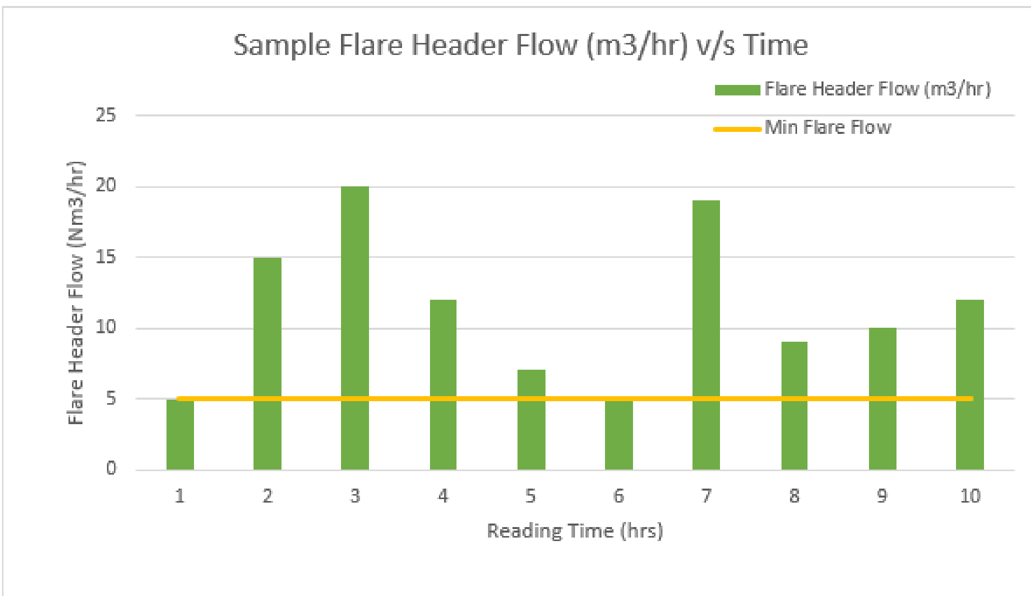


Figure 3 Sample Flare Gas Header Flow

Figure 3 shows a sample flare flow from a plant with a minimum of 5 Nm3/hr. of flare gas present in the header for 10 hours of operation. Assuming the flare gas is energy rich, flare gas recovery techniques can be used to utilize the excess amount of flare gas as a fuel or feed to other process plants.

The purpose of this study is to assess the operational risk in utilizing flare gas as a supplementary fuel to an existing bottom fired boiler to generate steam for power generation and provide additional heat to other process plants.

1.1. Motivation

In August, 2002, Global Gas Flaring Reduction partnership and “Zero Routine Flaring by 2030” initiative was launched by World Bank Group to curb flare gas emissions. 14 countries, 10 major oil and gas companies showed interest in overcoming the challenges for efficient utilization of flare gas (Brief, 2006).

SaskPower, a Canadian electric company has designed a flare gas power generation program, in which oil and gas companies can produce electricity through flare gas and sell electricity to SaskPower. TERIC Power Limited, another Canadian power company utilizes flare gas from oil wells to produce electricity. General Electric is producing electricity from associated petroleum gas in Russia. Pacific Ethanol, American bio-refining company is involved in utilizing waste gas to produce power.

Though in its initial stages, industries and international organizations has shown interest in utilizing flare gas as supplemental fuel. However, flaring during process upsets, emergency operations, equipment malfunctions, feed/product off-specifications, etc. are unpredictable event with varying gas flowrates, composition, pressure, flow, etc. In 2007, a fire tube steam boiler at Dana Corporation, Tennessee, exploded due to overheating of tubes (plant running in dry-fired state) injuring one person and damaging nearby facility (State of Tennessee Department of Labor and Workforce Development, 2007). Another incident took place in the same year at Dominion Energy, Massachusetts, when a coal fired steam boiler exploded due to tube failure. The incident caused 3 fatalities (Commonwealth of Massachusetts, Department of Public safety, 2008). From other literatures, improper fuel-air ratio, tube failure due to overheating and delayed ignition are some of the common causes of explosions (defined as combustion resulting in a rapid rise of pressure).

Thus, acknowledging the fact that flare gas utilization as a supplementary fuel has economic, energy and environmental benefit, it is also important to assess the operational risk on integrating flare gas containing high value combustible fuel to an existing boiler plant for cogeneration.

1.2. Problem Statement

When different flare streams with known variation in flowrate, frequency, composition, temperature from a specific plant is integrated to an existing boiler, it is desired to identify the technical operational risk during abnormal flaring. The study looks to address following:

- What are the hazards on integrating flare streams with an existing boiler?
- How the flare and the boiler process conditions affect the overall operation?
- What is the change in the probability of undesired events?

1.3. Overview of Flare Gas Recovery Techniques

Flare gas recovery systems were invented in 1970's, but companies started embracing these technologies by last decade. There are two approaches to flare gas recovery and re-usage.

I. Transporting flare gas through pipelines, reinjecting them into underground storage facilities, oil and gas reservoirs for carbon capture and sequestration (Abdollah Hajizadeh, 2017). LPG is produced by compressing associated gas, condensing heavier hydrocarbons and then separating the LPG. LNG is produced by refrigeration to nearly -162° C (after gas treatment) and shipped to tanks (Birnur Buzcu Guven, 2010).

II. Recover the energy stored in the waste or excess flare gas and utilize it as secondary fuel (e.g. fuel gas) in industry, or produce synthetic fuels through Gas to Liquid Fischer-Tropsch process, or produce electricity through cogeneration systems.

This study will focus on the second approach as the methods is at its incipient stage and needs further study. The general techniques used in flare recovery systems are following:

- Gas collection, compression and injection to fuel gas system.
- Gas-to-liquid (GTL) conversion.
- Fuel for Electricity Production.

1.3.1. Gas Collection, Compression and Injection to Fuel Gas System

The type of flare gas recovery technique to be used primarily depends on the flare gas composition and existing facility design. Due to the requirement of minimum modification to the existing plant setup, this process of flare gas collection from header, compression and reinjection to existing fuel gas system is preferred. Figure 4 shows a block diagram of flare gas compression, treatment and reentry to fuel gas system.

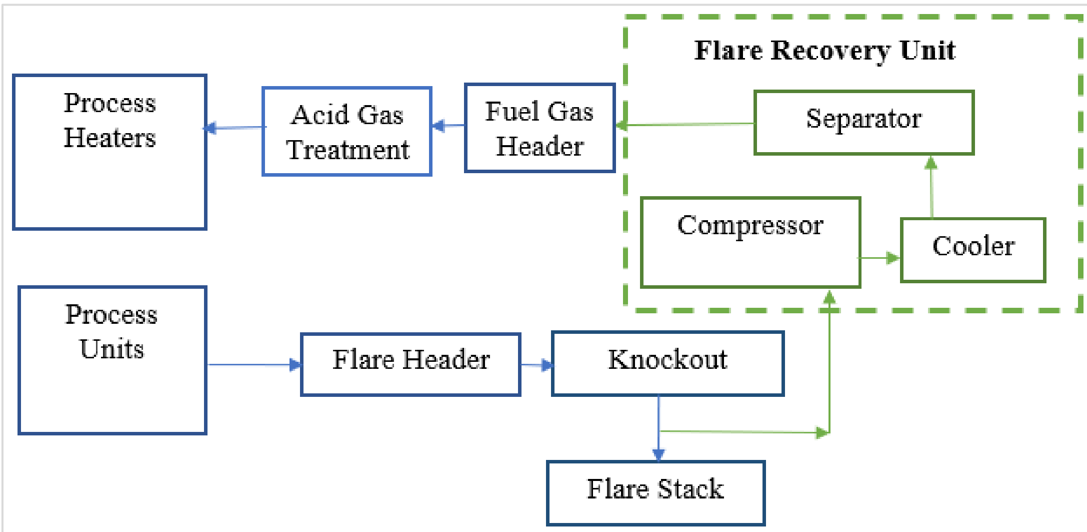


Figure 4 Diagram of Flare Gas Collection, Compression and Injection System

Generally, a flare gas is routed to Knockout Drum, which removes the carried out over liquid from the flare gas. Then the liquid free gas is sent to liquid seal (or water seal), which enables safer operation by providing back pressure in flare header. The flare recovery network is tied downstream of Flare Knockout Drum, operating simultaneously with flare stack operation. Typically, a minimum flare header pressure is maintained and the rest of the flare gas is sent through flare recovery network. If the flare gas flow is below the design flare stack flow, flare recovery system runs in recycle mode to maintain flare header pressure. Some of the fuel gas is also kept continuous through flare tip pilot lines. The common components present in compression and reinjection system are following:

- Compressor – Liquid ring compressor are the most common type of compression system used in flare recovery systems. Recent studies have also recommended using ejector systems. Compression design generally depends on the flare design conditions and intended flare recovery capacity. Liquid-ring compressors uses a sealing liquid (mostly water) in between impeller and compressor housing, preventing direct reversal of gas from compressor discharge. Also, liquid seal used as a sink to remove heat of compression and keeps the equipment in wet condition (P.W. Fisher, 2002).
- Water Cooler – After compressing the flare gas, the contents are cooled in an air/water cooler to remove heat of compression.
- Three Phase Separator – After cooling, Separator recovers the gases from the hydrocarbon condensate and sealing liquid/water. The sealing liquid is returned back to the liquid ring compressor. Gases are further routed to the fuel gas system (which are treated in acid gas treating facility for H_2S removal before using in process heaters, boilers, etc.).

Some of the plants using this recovery system are:

I. Valero Benicia Refinery in California, USA has a FGRS system designed to recover 5 MMSCFD of flare gas, using two reciprocating compressors (2 stage and 3 stage), and finally injecting recovered gas to meet refineries fuel gas system (total fuel gas demand in Benicia refinery is 75 MMSCFD). (Valero Refining Company - California, 2010)

II. Tesoro Martinez Refinery in California, USA has a FGRS system designed to recover maximum 5 MMSCFD of flare gas with two positive displacement compressors and routing recovered gas to sour fuel gas system for further treatment. The average flow to the refinery flare header was 0.8 MMSCFD in 2005 (Tesoro martinez Refinery, 2015).

1.3.2. Gas-to-Liquid (GTL) Conversion

Gas-to-Liquid technology is a very known process. The technology converts natural gas (containing methane) to valuable liquid fuels (e.g. gasoline, jet fuel, diesel).

As shown in the Figure 5, GTL process has three major steps:

- Natural Gas Reforming to produce synthesis gas ($CO+H_2$). Steam Methane reforming is preferred over partial oxidation, auto-thermal reforming.
- Catalytic conversion of synthesis gas to long chain heavier hydrocarbons using Fischer-Tropsch synthesis.
    - $\boldsymbol{(2n+1)H_2 + nCO \rightarrow C_nH_{2n+2} + nH_2O}$ ......Equation 1
- Hydrotreating of olefins and cracking of heavier hydrocarbons (or paraffins) to middle distillate (A. Ghorbani, 2012).

Figure 5 Block Diagram of a General GTL Process

Relatively low prices of natural gas to crude oil and carbon neutrality has boosted the GTL technology. Table 1 shows some of the examples of world scale and mini GTL plants.

Table 1 Example of Gas-to-Liquid Plants

Sl. No.	Type	GTL Plants	Gas Consumption (MMSCFD)
1.	World Scale	Atlas Methanol plant, Trinidad	260
2.	World Scale	Pearl, Qatar	1500
3.	Mini Scale	Juniper, Louisiana, United States	11

Every year 140 billion cubic meter of gas is flared. Around 500 million barrels of liquid fuels can be converted from this humongous amount of flared gas, worth 35 billion dollars (World Bank Group Energy and Extractives, 2015).

Flare gas flow less than 25 MMSCFD is considered as a good source for Mini-GTL. Compact GTL is the first company to build mini scale GTL plant with feed from associated flare gas. The GTL plant in Kazakhstan consumes nearly 25 MMSCFD of flare gas (World Bank Group Energy and Extractives, 2015).

The basic difficulties faced in converting flare gas to liquid fuels is the change in composition. However, GTL plants can process heavier feeds than methane with the addition of pre-reformer if the change is consistent for longer duration (World Bank Group Energy and Extractives, 2015).

1.3.3. Fuel for Electricity Production

A Combined Heat and Power (CHP) system, also known as cogeneration system, are systems which produce heat and electricity simultaneously. Generally, the electricity is generated by steam turbine or gas turbine and part of the steam generated is extracted as a heat medium for external uses. Cogeneration system utilizes the heat generated (along with the electricity) from the steam/gas turbine. The steam turbine efficiency of transforming fuel to power is 40% (Andrzej W. Ordys, 2009). Refer to Figure 6.

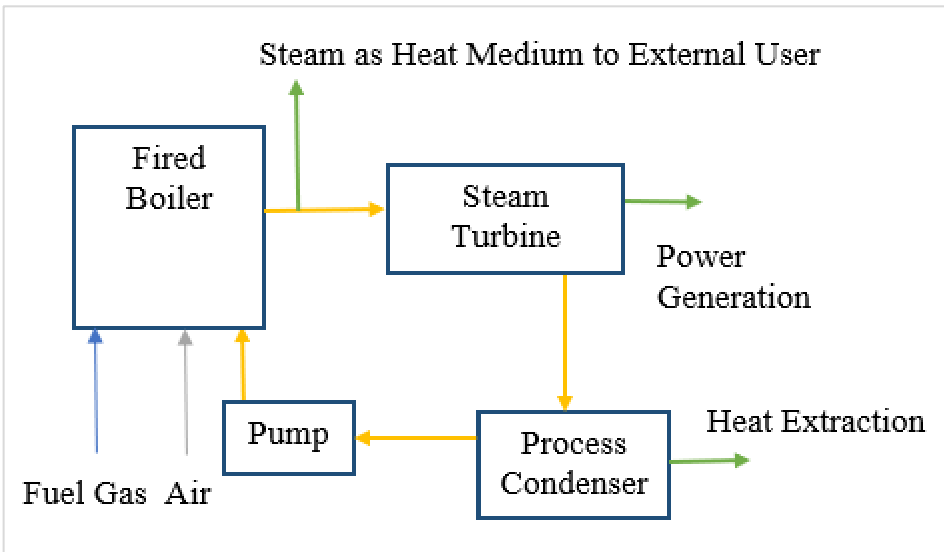


Figure 6 Cogeneration Plant with Steam Turbine

In Combined Cycle (CC) system, electricity is produced from gas turbine and steam turbine. Fuel gas is burned in gas turbine at higher temperature and the heat generated from exhaust gases of the turbine is utilized to generated steam, which again runs steam turbine to generate additional electricity. The total efficiency of combined cycle is 50% (A.W. Ordys, 1994).

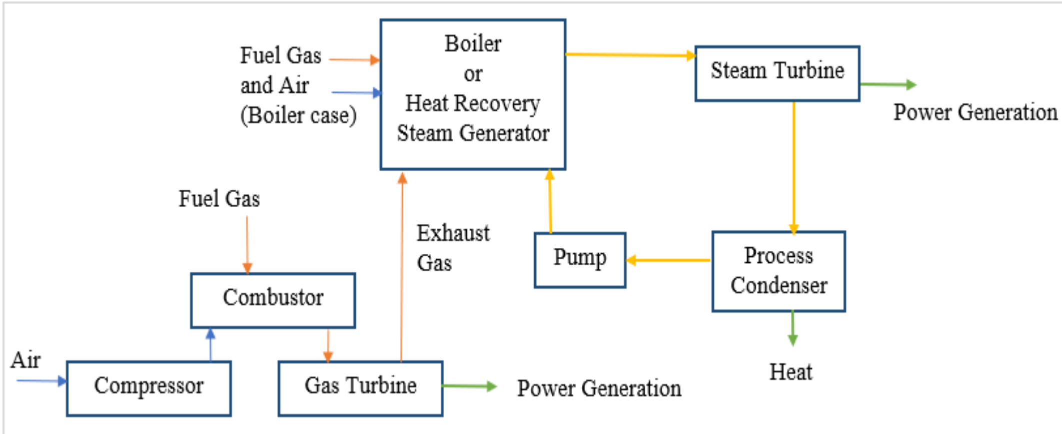


Figure 7 Standard Combined Cycle Power Generating System (A.W. Ordys, 1994)

The selection of prime movers (gas turbine, steam turbines or engine) depends on the plant fuel and power requirement. Gas engines offer partial loading and low emissions of gases due to clean burning and catalysis of natural gas. However, there is substantial energy loss due to partial load and lower electricity generation. Gas turbine transforms burning energy of the gas to rotational energy of its shaft. Steam turbines follow Rankin cycle, the high-pressure steam expands, rotating the turbine shaft to generate electricity. The operating temperature of steam turbines are lower than the gas turbines. Usually, gas turbines are used for higher power generation (1-300 MW per module) (Andrzej W. Ordys, 2009). Figure 7 shows a simple combined cycle power generating plant with a fuel gas fired boiler.

The use of flare gas in cogeneration systems is an important part of flare gas recovery technique. The flare gas (with considerable LHV) are routed to treatment unit (for H2S removal before utilizing the sweet flare gas in Gas Turbines or gas fired boiler for generating electricity (shown in Figure 8).

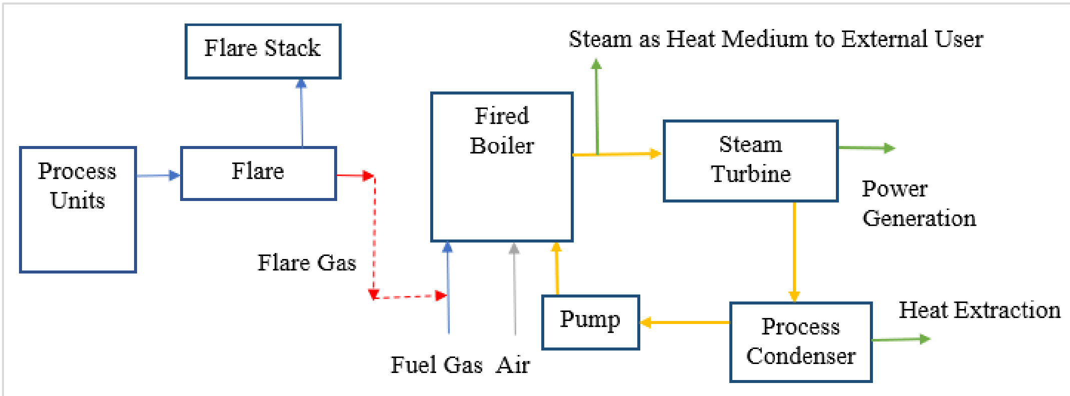


Figure 8 Combined Heat and Power Generation by Flare Gas Utilization

This cogeneration through flare gas helps in reducing GHG emissions, because, for the same amount of power produced, less amount of fresh fuel will be required.

1.3.4. Comparison of Flare Recovery Techniques

The application of flare gas recovery methods depends on the flare conditions, existing plant setup and user requirements.

Rahimpour et al. carried out two studies on different gas refineries with different flaring rate. First, a simulative study was carried out in Asaluyeh Gas Refinery to investigate the difference between each technique to recover 356.5 MMSCFD of flare gas containing 0.87 mole fraction of methane after sweetening. Table 2 shows the detailed values for comparing the different techniques (M.R. Rahimpour, 2012).

Result from the study are following:

- Gas compression technique has the lowest total capital investment with higher rate of return on investment as lesser equipment's are required.
- The annual profit is highest for cogeneration systems using gas turbine.

- For a downstream industry, with focus on increasing petroleum products while curbing CO_2 emissions, GTL process is a suitable alternative.

Table 2 Flare Recovery in Asaluyeh Gas Refinery (M.R. Rahimpour, 2012).

Sl. No.	Flare Recovery Techniques	Product Yield	Total Capital Investment (US MM Dollar)	Annual Profit (US MM Dollar)	Payback Period (Years)
1	Gas Compression	355.8 MMSCFD	72	151	0.5
2	Gas-To-Liquid	5,45,056 BPD	926	1155	0.8
3	Electricity Generation	2130 MW	1268	2746	4.76

Second study was carried out in Farashband Gas Refinery. The flare contains 4.176 MMSCFD of gas with 0.88 mole fraction of methane after sweetening. Table 3 shows the detailed values for comparing the different techniques in Farashband Gas Refinery (M R Rahimpour, 2012).

Result from this study are following:

- Gas compression has the lowest capital investment, however annual profit is lowest because the amount of recovered gas is low.
- Electricity generation with gas turbines, has the highest annual profits and lowest payback period (2.8 years) due to higher power generation.
- GTL has highest capital investment, but lesser carbon emissions as it utilizes CO_2 for fuel production.

Table 3 Flare Recovery in Farashband Gas Refinery (M R Rahimpour, 2012).

Sl. No.	Flare Recovery Techniques	Product Yield	Total Capital Investment (US Dollar)	Annual Profit (US Dollar)	Payback Period (Years)
1	Gas Compression	4.176 MMSCFD	3,360,000	1,225,510	2.8
2	Gas-To-Liquid	563 BPD	31,940,000	9,054,864	2.3
3	Electricity Generation	25 MW	33,355,084	14,053,600	3.3

Petri et al. carried out a techno-economic analysis of different flare gas recovery techniques. The study was to determine the utilization of 0.584 MMSCFD flare gas for-LNG production, LPG production or electricity generation. The gross heating value is 1162.43 Btu/SCF and contains 58 % mole percent methane and 12.9 % ethane primarily. The study concluded if the flare gas composition has considerable methane with a heating value between 950-1250 BTU/SCF and flare flow is moderate (<2.5 MMSCFD), power generation is favorable.

1.4. Applications of Flare Gas Recovery for Cogeneration – Current Status

The acceptance of flare gas for power generation is presently limited to gas fields (where associated gas is mostly flared) and renewable sources of energy.

SaskPower is one of the principal electricity supplying company in Sasktchewan, Canada. Coal industry has been largely used in Power generation. However, SaskPower relied on renewable sources of energy. The company has produced nearly 22% electricity from renewable energy from 2005 to 2014. In 2014, SaskPower designed Flare Gas Power

Generation Program, to support the development carbon neutral power generation techniques. As per the program, the oil and gas industries can sell electricity, generated from flare gas at a base rate $ 41.28/MWh in 2017 (Saskatchewan, 2015). In 2018, the project was revamped to Power Generation Partner Program (PGPP), allowing industries to sell as much as 25 MW of electricity, generated from flare gas (Electrical Line Magazine, 2018). TERIC Power Limited, an Alberta based power company, started using flare gas from Kerrobert oil wells owned by Sphere Energy, producing 750KW of electricity (Weyburn This Week, 2017).

In 2011, General Electric announced production of electricity generation from Nizhnevartovsk, Russia associated petroleum gas (a raw natural gas found with petroleum deposit) processing plant. GE will use natural gas from processing plant in combined cycle (Gas Turbine and Steam Turbine) to produce 400 MW of electricity, supplying to Nizhnevartovsk State Power Plant (General Electric, 2011).

Pacific Ethanol, a bio-refinery in California utilizes the waste gas to produce 3.5 MW power. The waste gases are introduced to a CHP gas turbine, containing power oxidizer (instead of a gas combustor, a technology provided by Ener-Core, a company based in California, United States) (Technologies, 2018). Ener-core claims the adaptability of the power oxidizer over reciprocating engines is in the utilizing lower LHV gases (below 30% methane) and producing lower quantities of NOx (Renewable Energy World, 2015).

2. PREVIOUS RESEARCH

Earlier research carried out is divided into four sections – flare gas recovery systems, hazards in heaters, risk assessment techniques and barrier management.

2.1. Flare Gas Recovery Systems

Cogeneration integration techniques have been studied by earlier research. Halwagi et al. has demonstrated the placement of heat engines and their integration with heat exchange networks (HEN). A power plant with a steam turbine is a heat engine, which following Rankin cycle where the boiler is an isobaric heating source with steam turbine and pump following isentropic expansion and compression and the condenser is constant pressure total condenser (El-Halwagi, 2017). Abdelhady et al. demonstrated the integration of solar energy and process heat for cogeneration systems (Faissal Abdelhady, 2015).

Flare gas recovery technology is in the industry from 1974 (MPR Industries, n.d.). Significant research has been carried out earlier on flare gas recovery techniques. Eman et. al. has provided an overview of gas flare recovery systems, discussing the techniques, application of the systems and role of Global Gas Flare Reduction Partnership (Emam, 2015).

Mourad et al. carried out simulation study for recovering flared associated gas containing high gas to oil ratio by compression and transferring the hydrocarbons to treating plant to valorize it as a fuel. The study showed requirement of multi-stage compressors and the relation between compressor energy and flare gas flow, pressure within each stage. The process indicated high initial investment due to high energy consumption and equipment requirement (Mourad Djebri, 2009).

Peeran et al. studied the use of special jet pumps instead of compressors to recover flare gas to recover and compress the gases to gas processing plants. (Syed M Peeran, 2015). Engarnevis et al. suggested use of gas ejectors for compression instead of conventional compressors, citing lower capital cost and operational cost. The results were conducted through mathematical modeling and the implementation was shown in Tehran oil refinery (A Engarnevis, 2013).

Rahimpour et al. carried out two studies on different gas refineries with different flaring rate to show the applicability of flare gas techniques. First, a simulative study was carried out in Asaluyeh Gas Refinery to recover 356.5 MMSCFD of flare gas containing 0.87 mole fraction of methane after sweetening. Second study was carried out in Farashband Gas Refinery to recover 4.176 MMSCFD of flared gas with 0.88 mole fraction of methane after sweetening. From the study, it was concluded that gas compression has the lowest capital investment and annual profit is highest for the cogeneration systems (M R Rahimpour, 2012) (M.R. Rahimpour, 2012).

Comodi et al. presented feasibility of flare gas recovery in a refinery. The study designed the recovery system using liquid ring compressor, using amine for gas treatment. The study showed economic feasibility for recovering 400 kg/hr. of flare gas (Gabriele Comodi, 2016).

Kamrava et al. proposed a process integration approach, identifying the flaring points from an ethylene plant, and re-introducing the flare gas as fuel source to a standby cogeneration system. The study assumed flare gas rates are steady and compared with non-integrated plants. The result showed integrated plants emitted lower CO_2 and higher annual operational cost savings (Serveh Kamrava K. J.-H., 2015).

Kazi et al. developed an optimization framework for a designing a cogeneration unit, recycling discontinuous flare gas from an existing ethylene plant as a fuel substitute for cogeneration unit. The study demonstrated the use of Pareto curve while deciding between two objectives – heating utility and power utility. The study suggested that the feasibility of recycling flare streams as a supplement fuel to reduce fresh fuel expenses. The heating value and Wobbe Index of the flare streams were sufficient to meet the power demand. However, the study did not consider the impact of variability in fuel quality (Monzure-Khoda Kazi F. M., 2015).

Another optimization framework was developed by M. Kazi, identifying different possible outcomes of integrating process plant, flare gas from normal and abnormal plant operation and waste water from treatment plants, using cogeneration and thermal membrane distillation (TMD) system. Referring to Figure 9, the framework considers the techno-economic and environmental factors and showed that the integration of process plant, waste water treatment and flare will provide higher power and heating utility with same energy consumption and additionally will have additional income (e.g. CO_2 savings) and lower payback back period (Monzure-Khoda Kazi F. E.-H., 2016).

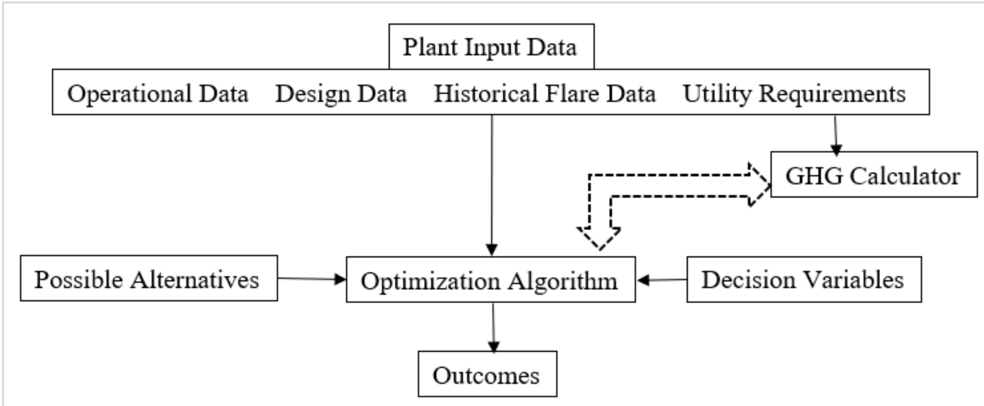


Figure 9 Multi-Objective Optimization Framework Integrating Process, Flare Gas and Water System (Monzure-Khoda Kazi F. E.-H., 2016)

Javier et al. proposed an optimization method for determining the optimal economic and environmental benefit of integrated system, utilizing flare from 3 different plants under normal and abnormal conditions to a cogeneration plant. Three alternatives were identified – total flare gas burning the flare stack, partly mixing flare gas as supplemental fuel to cogeneration unit, and mixing entire flare stream with fresh fuel to cogeneration system. The uncertainties in gas flows, heating value and natural gas price were included by choosing 50 different scenarios. After comparing, the result showed environmental and economic benefit for reutilizing flared streams. The study however, used random Wobbe Index values for calculation (Javier Tovar-Facio, 2017).

2.2. Hazards in Heaters

Dugue et al. reviewed the fired equipment safety. Starting from the NFPA-85 codes for fire heated boilers (till 2011), the work discussed common hazards, incidents and safeguards in fired heaters. The main hazards in heaters are – explosions and tube ruptures. Explosions occur after flame blow off (rich fuel condition), when gas is reignited or air is introduced swiftly. These can damage fired heater floors, bridge walls. Tube rupture occur due to loss of cooling medium or overheating of internal tubes. Large length/diameter and presence of partition wall can result in overpressure explosions (Dugue, Fired Equipment Safety in the Oil and Gas Industry, 2017).

Hawryluk et al. studied the hazards of fuel rich combustion in furnaces. The flue gas was analyzed to identify the richness of fuel. The temperature, residence time and mixing were assumed to be normal. The study demonstrated that a furnace running on excess natural gas, above 1298 °F, all the methane will be converted to hydrogen in the flue gas. This rich flue gas can pose an overpressure hazard, if they quickly mix with fresh air inside the furnace at elevated temperature (Hawryluk, 2008).

Ogle et al. investigated the fire-tube explosion due to detonation of natural gas. The methodology included inspection of incidents, analysis of damage patterns, comparison of incident scenarios, finally determining the cause. From a boiler shell explosion incident, the study identified that firebox flameout led to accumulation of unburnt natural gas-air mixture, which reached the stack and ignited. The combustion wave propagated back, accelerated and detonated (Russell A. Ogle, 1999).

Sharifi et al. investigated the use of hydrogen along with regular fuels in furnace. The study involved analyzing the various hydrogen properties- energy content, energy density, radiant heat, diffusivity and flammability. The result showed environmental emission reduction but more propensity towards critical operation due to hydrogens high combustibility (Vahid Sharifi, 2012).

Ditaranto et al. studied the effect of hydrogen combustion on refinery fired heaters. Computational Fluid Dynamics simulations were carried while switching refinery fuel (methane) to hydrogen. The radiative heat load change was observed to be nominal. Burner overheating and higher NOx was observed during direct change from methane to hydrogen due to high flame temperature and flame speed. Modifications in burner resolved overheating problems (M. Ditaranto, 2013).

Jones et al. studied the addition of hydrogen in natural gas appliances. The research evaluated the heating value (Wobbe Index) with higher hydrogen composition and studied the flame stability and flashback phenomenon in natural gas appliances. The study suggested (considering the port size of burners are relatively larger for fixed volumetric flow) hydrogen addition up to 34.7 mole% instead of 10 mole% hydrogen based on the minimal Wobbe Index limit of 49.75 MJ Nm^{-3} on natural gas appliances (Daniel R. Jones, 2017).

Lowe et al. conducted a simulation test of fired heater while changing the fuel from natural gas to hydrogen. The study checked heater conditions, NOx burner performance and fuel gas system reliability. The burners in the existing facility was replaced with NOx burners due to hydrogen's higher flame speed (10 times that of methane) and flame temperature. The study concluded that the hydrogen will increase the radiation tube temperature, arch temperature, while reducing the convection section heat. Moreover, the fuel gas header showed a rise in pressure drop across mixing drum (Cliff Lowe, Nick Brancaccio, Dan Batten, Chris Leung, 2011).

Ramirez et al. investigated a firebox explosion in a steam boiler. The systematic root cause analysis method was used which primarily focused on Burner Management System (BMS). The failure of the boiler was identified due to multiple causes – indication failure of combustion control and failure of low air sensor led to failure of emergency shutdown. Moreover, closure of forced draft fans, damper resulted in over-pressurization with rich fuel (NFPA 85 recommends minimum 25% opening for air passage) (Juan C. Ramirez, 2010).

Vries et. al. carried out test on boilers in Denmark, having lean burning (fully premixed type, where air and air are combined upstream of the burner) tendency. When a boiler operates at constant fuel gas pressure and constant air flow with H2 above 70%-80%, due to hydrogen high flame velocity, flashback might result, resulting in boiler shutdown (De Vries Harmen, 2007).

Bhangu et al. applied probabilistic risk assessment to analyze reliability of power plant. The study used Pareto method to identify critical failures (contributing to 85% of unit shutdown) and then applied fault tree analysis to identify failure of the plant (N. Singh Bhangu, 2015).

Saleh et al. studied the reliability of steam boiler including operational, maintenance aspects. The hazard identification was carried out through fault tree analysis, which identified critical component failure rates. A visual basic language was used to evaluate reliability with time. The study concluded failure rate was maximum at the early stages (in seven month) and then follows a constant rate (Faik Lateef Saleh, 2012).

2.3. Risk Assessment Techniques

Hazard identification is the first step of risk assessment. The hazard identification is qualitative and the methods chosen are based on the organization, process facility, their size, operating data and phase of the plant. Hazard identification procedures are generally classified into non-scenario based (created on experience) and scenario based (predictive and analytical). Safety review, checklist analysis, preliminary hazard analysis, relative ranking indices, checklist analysis are non-scenario based whereas What-If, HAZOP, Fault Tree, Event Tree, Bow-tie analysis, Failure Modes and Effect analysis are scenario-based hazard evaluation methods (Safety, 2008).

Center for Chemical Process Safety mentions that for a regular operation, all but, relative ranking and preliminary hazard analysis are used (they are more used during initial design phase of the process). For single failure analysis, FMEA, What-If are preferred over Fault Tree Analysis and Event Tree Analysis. When previous information and experience is limited, prediction-based HAZOP or Fault Tree Analysis is used. However, hazard evaluations are qualitative method and is based on collective team knowledge, experience and information (Safety, 2008).

Often the hazard evaluation team can miss scenarios due to the overwhelming plant complexity, organizational factors. Incidents can happen after year of operation without any advance warning signals. The interaction of physical and chemical properties and material behavior cannot be precisely predicted. If the predicting of systems is not feasible, observing and learning from the process is of paramount importance. (Hans J. Pasman, 2017). Center for Chemical Process safety has identified some of the potential limitations of hazard identification – completeness in identifying all hazards, reproducibility, inscrutable, competency of team and subjectivity of analysis (Safety, 2008).

Paltrinieri et al. defines atypical incident scenarios when identified scenarios from hazard evaluation deviate from normal prediction. This happens because of limited information or professional’s unawareness to warning signals or related events. Since the rarity of the events, atypical incidents are categorized as as unknown unknowns (unaware of events and unavailability of related information) and unknown knowns (unaware that we can learn from related information available). For example, a vapor cloud explosion in Buncefield oil depot, 2005 is an example of atypical incident, where the earlier hazard identification showed to be gasoline pool fire as a credible loss control event (Nicola Paltrinieri, 2016).

Bobbio et al. studied the use of fault tree in Bayesian Network to evaluate the probability of dependable systems. The method was tried on hardware and software systems and identified that both forward (prediction) and backward (diagnosis) analysis can be carried out in BN. Moreover, the dependencies can be modeled in BN while considering uncertainties in failure rates (A. Bobbio, 2001).

Khakzad et al. carried out the comparison between fault tree and Bayesian Network. The study concluded the advantage of using Bayesian Network over fault tree due to the inclusion of uncertainty, multiple dependencies, updating prior probabilities of components and the ability to incorporate different state for a single variable (Nima

Khakzad, Safety Analysis in Process Facilities: Comparison of Fault Tree and Bayesian Network Approaches, 2011).

Bearfield et at. mapped event tree to Bayesian Network, demonstrating the relationship of events, where events frequency is conditioned on prior event occurrence. With a derailment case study, the study represented the influence of causal on the probability of consequence (George Bearfield, 2005).

Unnikrishnan et al. applied Bayesian Network to event tree. The prior probabilities of top event and barriers are applied to find the conditional probabilities of consequence from top event. The study demonstrated the usefulness of Bayesian Network in predicting consequence, diagnosis of causal factor influencing consequence, and updating component information to the methods (G. Unnikrshnan, 2014).

Risk assessment is generally divided into three sections – dynamic risk assessment, operational risk assessment and real time risk assessment. Risk analysis is to identify and estimate risk level. The concepts of dynamic, real-time and operational risk assessment are intertwined and used interchangeably. Figure 10 shows subtle differences between each risk assessment methods.

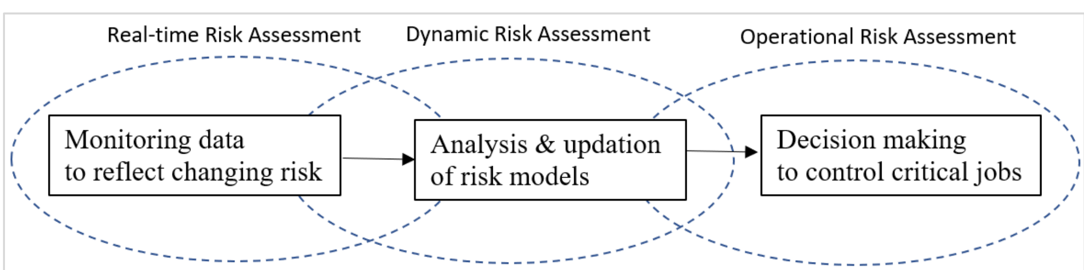


Figure 10 Risk Assessment Techniques (Xue Yang, 2018)

Real time risk assessment is evaluating the risk, when real time operation sate is added to risk assessment models. Data collection can be through process analyzers (e.g. H2S values) or management systems (e.g. near miss reports).

Quantitative Risk Assessment is a technique to identify and evaluate overall risk. However, it cannot identify dynamic changes in the plant. The data incorporated in QRA are limited, old and does not address uncertainties. Dynamic risk assessment purpose is to update calculated risk which includes addition of new data to evaluate the ever-changing plant conditions, such as evaluating dynamic prevention and protection safeguards, consequences. Importantly dynamic risk assessment considers causal interaction or dependency of components. The dependency of causal factors, top events, safeguards, consequences are modeled in Bow-tie diagram or Bayesian Network (Xue Yang, 2018).

Operational risk assessment is to identify, analyze and evaluate the overall risk in in a critical operational event. In operational plant, decisions are divided into four categories –

- Risk related with human and organizational errors, such as filling column above design limit (e.g. splitter column overfilling).
- Incident risk while operating a facility/plant, such as fire in the facility.
- Risk associated with critical tasks such as handling plant emergency.

Yang et al. developed a framework for operational risk assessment, where a Bayesian Network is used to calculate the probability of an event using precursor events and loss functions are used to show time-dependent consequences in terms of monetary values. The BN and loss functions were integrated to develop a real time risk trend (Ming Yang, 2015).

Khakzad et al. carried out a quantitative risk assessment of drilling operations using Bayesian Network and Bow-tie method. Initially, the accident scenarios are developed in event tree and fault tree and then translated to Bow-tie model. Second the Bayesian Network was modelled for accidents. On comparison, it was concluded from the study that Bayesian Network can provide the dynamic analysis of drilling operation as it includes common cause failures and explicitly analyses the conditional dependencies, while learning from precursor events (Nima Khakzad, Quantitative Risk Analysis of Offshore Drilling Operations: A Bayesian Approach, 2013).

Abimola et al. applied Bow-tie model (graphical representation of cause and consequence) and real time safeguards failure (blowout preventer) probability to assess the dynamic risk of offshore drilling operation. The prior probabilities of components failure are updated in Bow-tie model (Majeed Abimbola, 2014).

Barua et al. showed real time cause effect relationship by dynamic operational risk assessment. The method demonstrated dependency of variables and the effect of change on dynamic Bayesian Network. The study developed a dynamic fault tree and mapped it onto dynamic Bayesian Network, to identify and evaluate operational changes in components (Shubharthi Barua, 2016).

Pakorn et al. carried out probabilistic risk assessment by applying Bayesian Network to risk assessment techniques. HAZOP and Bowtie were mapped in continuous type Bayesian Network. The 11 process safety indicators data were collected from IOGP for a year, represented into prior distribution, assimilated into three categories – mechanical, operational and personnel integrity and finally incorporated into Bow-tie-BN and HAZOP-BN model. Methodology shown in Figure 11 (Pakorn Chaiwat, 2016).

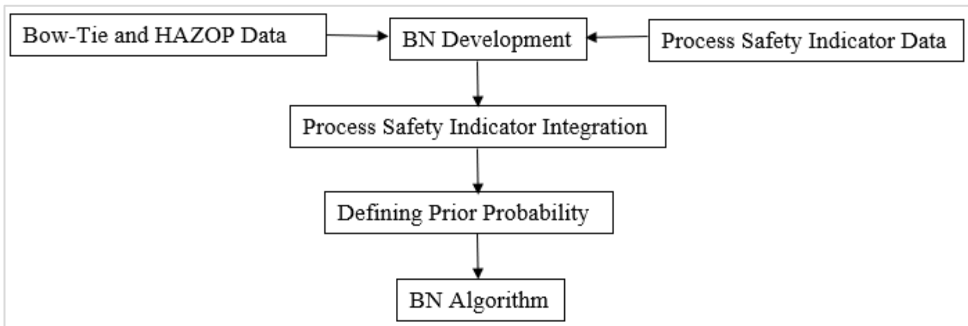


Figure 11 Integration of Bayesian Network and Process Safety Indicators in HAZOP and Bow-Tie Study (Pakorn Chaiwat, 2016)

Susana et al. proposed a framework to identity hydrates in offshore drilling operations. She applied Bayesian Network from Bowtie method and compared the results with kinetic and thermodynamic models with sensitivity analysis (Susana Leon Caceres, 2017).

2.4. Barrier Management

Barriers are physical or non-physical component that prevent, control, or mitigate undesired event. A barrier can be classified into technical (e.g. sensors), operational (e.g. operating pump) or organizational (e.g. procedures). Technical barriers are subdivided into Safety Instrumented Systems (SIS), safety systems without internal instrumentation logic, and risk reduction systems (e.g. egress components) Performance Influencing Factors (PIF) or Risk Influencing Factors (RIF) are helps the functionality of barriers (e.g. preventive maintenance). Further, the barriers are divided into proactive and reactive barriers based on their preventive or mitigating roles. Johansen et. al. discusses the uses and reproduces the challenges in barrier management systems (Inger Lise Johansen, 2015).

Haddon et al. proposed a comprehensive application in cause-consequence model, which he called Hazard-Barrier-Target method. He proposed ten strategies to reduce hum and other losses through barriers (Haddon, 1973).

Sklet et al. carried out a thorough review of the barriers, classifying the barriers and providing further information about the performance of safety barriers (Sklet, 2006). The ARAMIS project under SEVESO-II directive, classified the barriers based on their action as – avoid, prevent, control, protect. Table 4 represents classification of barriers based on the action proposed by Duijm et al.

Table 4 Safety Barriers Classification (Sklet, 2006)

Parameters			*Accident Sequence*		
Process Condition	Normal		Initial	Concluding	Injury
Barrier Actions	Avoid	Prevent	Control		Protect

2.5. Limitations of Current Research

Studies have been carried out on techno-economic analysis of flare gas integration to cogeneration systems. However, limited research has been done to address the process risks involved in using inconsistent flare gas from a plant to an existing fired boiler for cogeneration. Some of the process risk unaddressed in previous research are:

A. Effect of flare gas quality on cogeneration system.

In an abnormal flaring, flare gas will have inconsistent quality and process conditions. Each composition will have a different impact on the cogeneration performance.

- If flare gas has hydrogen, then as hydrogen is carbon free, it produces flameless combustion and is not visible to naked eye. Moreover, hydrogen has lower heating value of 51,585 Btu/lb. compared to 20,267 Btu/lb. of natural gas. Also, hydrogen has higher laminar flame speed (7.3 times more than methane in atmospheric conditions). These reasons can lead to hydrogen fuel burning fast, chances of flashback and higher temperature near burners and refractory.
- Condensate carryover along with the fuel in boiler combustion chamber can plug the burners and burner inlet strainers, resulting in reduced load operation (assuming there are limited standby burners available). Water carryover with fuel cause pressure shock due to the expansion of water to steam at high boiler temperature (expansion factor is 1600 times), which can lead to tube and burner failure. Additionally, excess water retention can lead to equipment corrosion and burner flame put off.
- Flame impingement in a boiler is due to improper air fuel mixing can raise the tube metal temperature. Un-uniform air fuel can cause flameout and instant startup will lead to firebox explosion.

B. Risk assessment methods

- Hazard identification is generally qualitative which is based on experience or prediction. The evaluations are sensitive to safety professional's judgement, risk acceptance and assumptions. Thus, for same information, different professional will have different outcomes, which lead to subjectivity. Moreover, if the teams experience and competency is less, then the outcome of the evaluation can be less effective.

- For new process technologies or new operational changes, the availability of information on related incidents, warning signals are limited. Then predictions are used in place of experience. These prediction based hazard identification can be incomplete and may miss some low occurrence, high probability events.

3. PROPOSED METHODOLOGY

Hazard evaluation is a systematic way to identify and analyze the significance of hazardous situation in a process. Hazard evaluation is extensively used for checking the occurrence of undesired events in design and operation by hazard identification, scenario development and assessment of risk. Considering the previous research limitations, this section proposes a framework for assessing the operational risk in managing flare gas for cogeneration (shown in Figure 12).

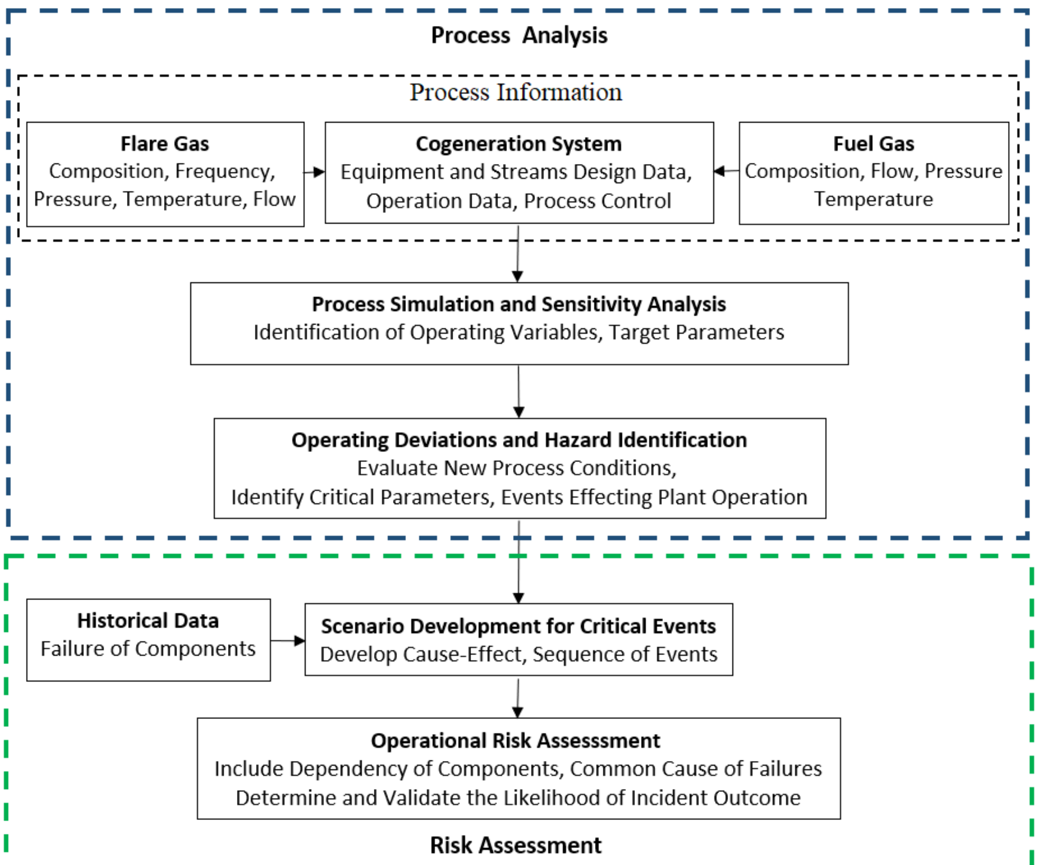


Figure 12 Framework to Assess Operational Risk of Using Flare Gas for Cogeneration

3.1. Process Analysis

To assess the hazards and risk in a process, it is essential to define the process boundary first. After the process boundary is selected, following steps are followed- collection of data, building of process model, checking the response of operating parameters through sensitivity analysis.

3.1.1. Process Information

After defining the process boundary, design and operational data are collected for the particular section of the plant from process safety information, which includes Process Flow Diagrams (PFD), Process Flow and Instrumentation Diagrams (P&ID), process control schemes, process package data on equipment's, process streams, utility streams, instrument device, preventive and mitigating safeguards.

3.1.2. Process Simulation and Sensitivity Analysis

After the process design data and operational data are collected, process simulation is carried out to predict the response of a defined system under the given set of conditions (Dominic Foo, 2017). The general steps followed for process simulation are shown below:

- Providing Information – Choose chemical components and thermodynamic model. All the pure light gas components are entered and Peng-Robinson equation of state (EOS) is chosen because for combustion application with wide range of change in temperature and pressure PR-BM is preferred (Aspen technology Inc., 2000).
- Building Flowsheet - Choose model, join individual streams.

Simulation environment has main flowsheet. In this the sequence of operation will be defined. The material steams will be chosen and the conditions will be defined based on the collected data. Sequentially other equipment's and energy streams will be defined and finally connected together.

- Running Simulation – Give specification, converge data and validate simulation results.

After providing the data and specifying the streams and equipment and simulation can be run. In the workbook section, all the conditions of material, energy streams and unit operations can be monitored and validated.

Sensitivity analysis is carried out to evaluate the effect of manipulated variable on the target parameter or on the overall process. For example, in this study the fuel quality was considered as one of the manipulated variables and the effect on boiler was observed (Aspen technology Inc., 2000).

After sensitivity analysis, the manipulated variables which affect the process condition and are not consistent with the operating philosophy and/or design limit are identified as process hazards (also can be referred as primary events). The change in process operating condition are classified into intermediate events. Thus, from sensitivity analysis, the critical parameters and their effect are identified.

It is important to note that through simulation and sensitivity analysis, only technical safety items (physical parameters, e.g. pressure, temperature, etc.) will be identified. The human factors will be identified during scenario development.

3.2. Operational Risk Assessment

After the identification of critical parameters and the type of effect on the system, a hazard scenario will be developed to evaluate the occurrence of loss control event and the probability of occurrence of various consequences.

3.2.1. Scenario Development

Fault tree and event tree will be used for hazard scenario development. Fault tree and event tree are probabilistic methods to identify and analyze loss of control event (or top event). The top event is when control is lost, and is the common node between cause and consequence diagram. The top event is defined as an event which can lead to many consequences, such as tube rupture, firebox explosion.

Fault tree is a deductive approach, finding causes which can lead to top event. Fault tree uses Boolean logic (two possible values). All the events in the fault tree diagram will receive a probability of occurrence, based on which top event occurrence will be predicted.

Following steps are followed for construction fault tree diagram:

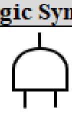
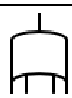
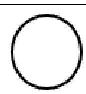
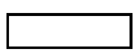
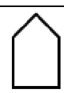
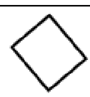
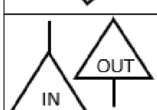
- Define top event (loss of control).
- Construct contributing causes/events leading to the top event.
- Calculate the frequency of top event.
- Identify the minimal cut sets.

Several logic functions are used for fault tree are shown in Table 5. The fault tree diagram is prepared using the logic functions. Then the failure probability of all the components are provided. The top event frequency calculation can be done in two ways-

I. From fault tree diagram – calculations are carried out across all logic gates with probability of failure updated at each event. AND gate multiplies the failure probabilities whereas OR gate multiplies the reliability of components.

II. From minimal cut sets- Cut sets are the exclusive pathways followed by the base events to reach top event. Thus, from the diagram, minimal cut sets (or the failure modes) are identified. Usually for large events with higher failure probabilities, the estimation of failure probability from minimal cut sets are more conservative than through fault tree diagrams (Joseph F. Louvar, 2011).

Table 5 Logic Functions for Fault Tree (Joseph F. Louvar, 2011)

Logic Symbol	Name	Description
	AND	Output depends on simultaneous occurrence of input event
	OR	Output depends om occurrence of any of the input event
	BASE EVENT	The starting event
	INTERMEDIATE EVENT	Formed in between events
	EXTERNAL EVENT	Event which is boundary condition in the diagram
	UNDEVELOPED EVENT	Event cannot be further developed
	TRANSFER SYMBOL	To transfer diagram to other sheet

Fault tree is complemented by Event tree. Event tree also follows Boolean logic. Following steps are followed for constructing event tree diagram:

- Identifying top event (same as fault tree).
- Construct the event sequence in the even tree diagram.
- Identifying safeguards or barriers to mitigate the consequences.
- Calculate the frequency of consequences.

Event tree is an inductive approach, starting with a top event and working toward worst case scenario. Top event is first written and mitigating barriers are incorporated with either fail or succeed probabilities.

Barriers are physical or non-physical component that prevent, control, or mitigate undesired event. A barrier can be classified into technical (e.g. sensors), operational (e.g. operating pump) or organizational (e.g. procedures). Technical barriers are subdivided into Safety Instrumented Systems (SIS), safety systems without internal instrumentation logic, and risk reduction systems (e.g. building exits) (Inger Lise Johansen, 2015).

The calculation at each junction is carried out from the failure rate of safeguards and probability of top event as shown in Figure 13 (Joseph F. Louvar, 2011).

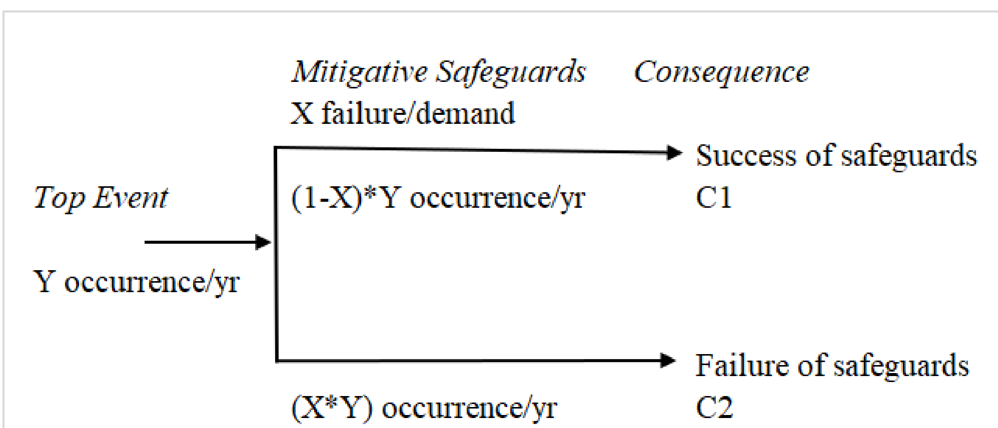


Figure 13 Sample Event Tree Analysis (Joseph F. Louvar, 2011)

3.2.2. Failure Data Collection

There are several reliability databases which are used by academia and industry. Offshore and onshore reliability database (OREDA) and Lees loss prevention in the process industries are primarily used for the failure rates of equipment's, instrument's and human error. From OREDA database, individual components are assumed to be operating in homogeneous operating conditions and the plant is operating in useful life phase with constant failure rate. The probability of failure is expressed using Poisson distribution (Joseph F. Louvar, 2011).

$$\boldsymbol{R(t) = e^{-\mu t}} \text{......Equation 2}$$

$$\boldsymbol{P(t) = 1 - e^{-\mu t}} \text{......Equation 3}$$

where, R(t) = Reliability of component in time, t,

P(t) = Probability of component failure in time, t,

μ = constant failure rate of component.

3.2.3. Bayesian Network for Qualitative Risk Assessment

The ability of bi-directional interpretation along with probabilistic calculation gave rise to Bayesian Network in 1970's. Bayesian Network derived from Bayes' theorem, which encompasses conditional probability.

$$\boldsymbol{P(hi|E) = \frac{P(E|hi) \times P(hi)}{P(E)}} \text{......Equation 4}$$

$$\boldsymbol{P(E) = P(E|h1)P(h1) + P(E|h2)P(h2) + \cdots . P(E|hn)}$$

$$= \sum_{i}^{n} \boldsymbol{P}(\boldsymbol{E}|\boldsymbol{hi})\boldsymbol{P}(\boldsymbol{hi}) \text{……Equation 5}$$

where, P(H) = Probability of hypothesis, H,

P(E) = Probability of evidence, E,

P(H|E) = Probability of hypothesis, H given evidence, E.

A Bayesian Network is a representation of the dependency between variables and helps in understanding the propagation of effects. Bayesian Network pictorially represents nodes, arcs and Node Probability Tables (NPT). Nodes represent the variables, arcs elicit the dependencies between the variables and NPT displays the probability distribution. There are certain advantages of mapping fault tree and event tree on Bayesian Network.

- Bayesian Network are bi-directional, i.e. forward and backward analysis can be carried out.
- Bayesian Network are more explicit, which can have more than two states whereas the fault tree and event tree has only two states. For example, in Bayesian Network flow node can be mentioned as high, low and medium.
- In Bayesian Network, the parent nodes (primary components) can have dependency (e.g. running pump and standby pump failure due to dependency on power). In Fault tree basic events are independent.
- Event tree lacks the ability to demonstrate individual events influence on top event (backward analysis). Moreover, the states are only binary, rendering less clarity.
- Bayesian Network calculations are more accurate than fault tree analysis. In discrete mode, while calculating the event occurrence in fault tree through cut sets,

the intersection of multiple events is dropped $P(X \cap Y \cap Z)$ (Norman Fenton, 2013)

$$\boldsymbol{P}\ (\boldsymbol{X} \cup \boldsymbol{Y}\boldsymbol{Z} \cup) = \boldsymbol{P}(\boldsymbol{X}) + \boldsymbol{P}(\boldsymbol{Y}) + \boldsymbol{P}(\boldsymbol{Z}) - \boldsymbol{P}(\boldsymbol{X} \cap \boldsymbol{Y}) - \boldsymbol{P}(\boldsymbol{Y} \cap \boldsymbol{Z}) - \boldsymbol{P}(\boldsymbol{Z} \cap \boldsymbol{X}) + \boldsymbol{P}(\boldsymbol{X} \cap \boldsymbol{Y} \cap \boldsymbol{Z}) \ldots\ldots \text{Equation 6}$$

Steps followed in mapping fault tree and event tree to Bayesian Network are:

I. Fault tree to Bayesian Network – The primary events, intermediate events and top events from fault are selected as root nodes, intermediate nodes and leaf node in Bayesian Network. The root nodes are provided with prior probabilities from OREDA database. The intermediate and the leaf nodes are defined with conditional probability table (Nima Khakzad, Dynamic Safety Analysis of Process Systems by Mapping Bow-tie into Bayesian Network, 2013).

II. Event tree to Bayesian Network – The safeguards are shown as nodes with binomial states (e.g. fail or success), with prior probability values from OREDA database. The safeguard nodes will be connected to each other if they are dependent on each other's response. The safeguard node (e.g. alarm) will be connected to consequence node (e.g. tube rupture), if the consequence states are influenced by failure/success of safeguards. The consequence node in Bayesian Network has same number of states as the consequences in the event tree. Conditional probability table is assigned to consequence node.

After the construction of Bayesian Network for the identified scenarios, quantitative inferences can be made on the occurrence of various consequences, the influence of critical parameters on top event and comparison of process conditions with different operating parameters.

3.3. Software

Aspen Hysys V10 software is used for process simulation and sensitivity analysis. Peng-Robinson-Boston Mathias equation of state is used for the cogeneration system due to its acceptability in natural gas combustion process and wide range of temperature and pressure application (Maria Jesus Guerra, Aspen Process Engineering Webinar, 2006).

AgenaRisk 6.0 software is used for constructing Bayesian Network. The software is capable to run both discrete and continuous variables and has the ability to do sensitivity analysis through tornado graph. It can be used for prediction, diagnosis as well as for causal factor and dependency evaluation (Agena Risk Inc, 2012).

4. CASE STUDY

The operational risk assessment flowsheet to identify the process hazards, their associated effects on operation and quantifying the occurrence of untoward events is shown in Figure 14.

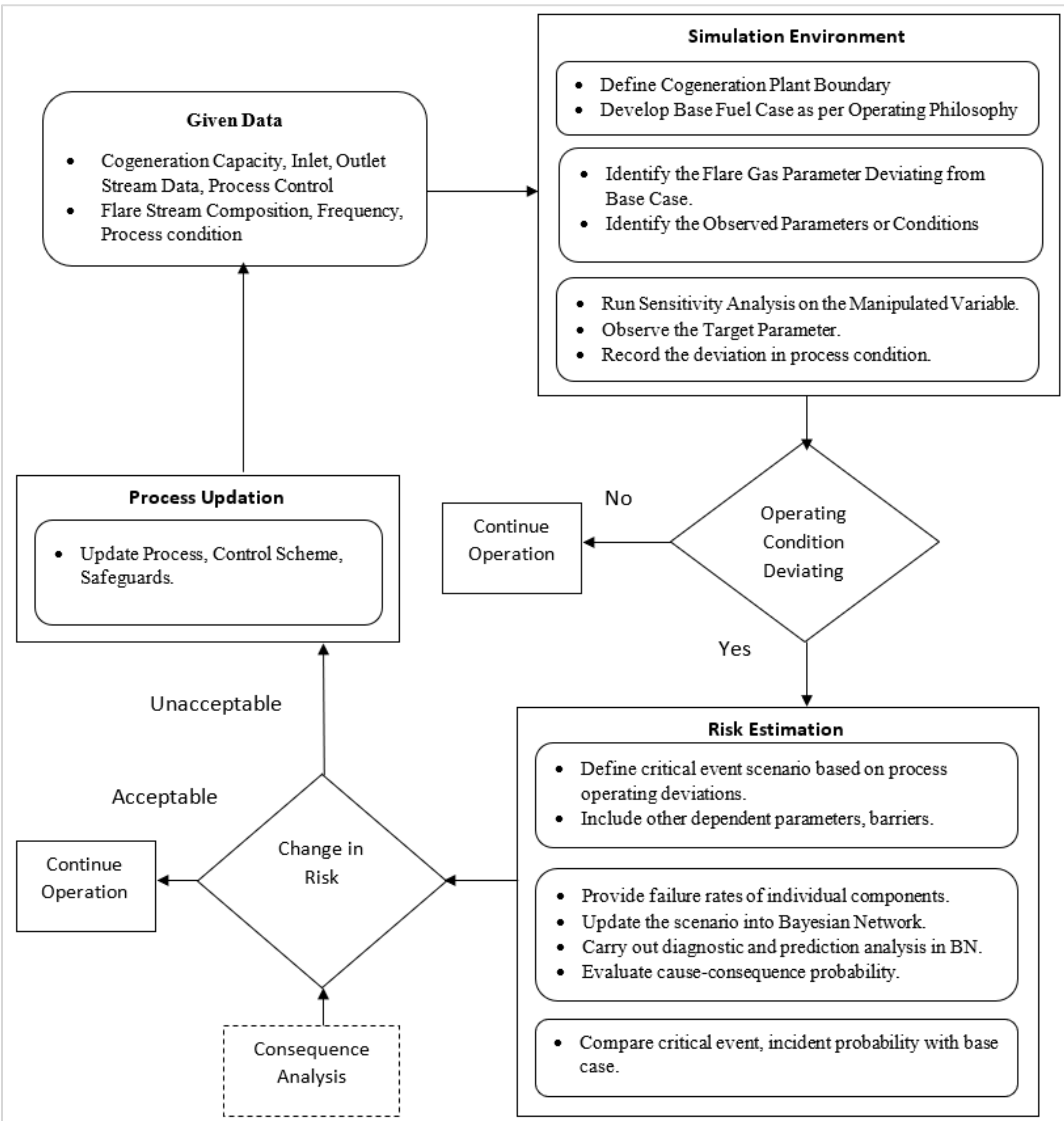


Figure 14 Flowchart of Operational Risk Assessment of Boiler System

4.1. Process Description

In this case study, a steam boiler with a bottom fired furnace and a natural draft circulation is chosen. Natural gas is used as a fuel to the furnace in base case study. Fuel gas header pressure is maintained at 50 psia and flow is controlled through controller, FIC-1. Before natural gas enters the firebox, pressure is maintained through pressure regulator, PRV-1. In natural gas case, natural gas flows at 2866 lb/hr at 14.5 psia and 108 °F to the firebox. Dry air is controlled through flow controller, FIC-2 at 49,950 lb/hr and reaches fired boiler at 392 °F. The air to fuel ratio is maintained at 9.92 (mole basis).

Boiler feed water at 37,600 lb/hr at 290 psia first exchanges heat with flue gas in economizer and then is routed to radiation and convection zone of furnace. The diversion of boiler feed water to radiation zone is 68.5% as most of heat duty of furnace is from radiation zone. The radiation outlet steam temperature is 429 °F and convection zone outlet steam temperature is 435 °F at 275.6 psia. The steam produced are operated at 20 °F above steam saturation temperature (at 275.6 psia, steam saturation temperature is 410 °F). Both the steam mixes in the Steam Drum-2 and then goes to steam turbine for power generation.

The flue gas from convection section goes to economizer exchanging heat with inlet boiler feed water (5 MM Btu/hr duty). Additional heat from the flue gas is provided to the external process at 14 MM Btu/hr. Before routing flue gas to atmosphere, it exchanges the residual heat with air in air-preheater (3.5 MM Btu/hr). Flue gas at stack outlet to atmosphere is maintained above 320 °F, to avoid corrosion due to Sulphur condensation, if any. The total heat duty of furnace is 43.3 MMBTU (excluding heat provided to external process). The efficiency of heater is assumed to be 40%.

Figure 15 and Table 6 shows the initial condition of boiler to meet 37,600 lb/hr of steam with boiler duty less than 60 MM Btu/hr (Serveh Kamrava K. J.-H., 2015).

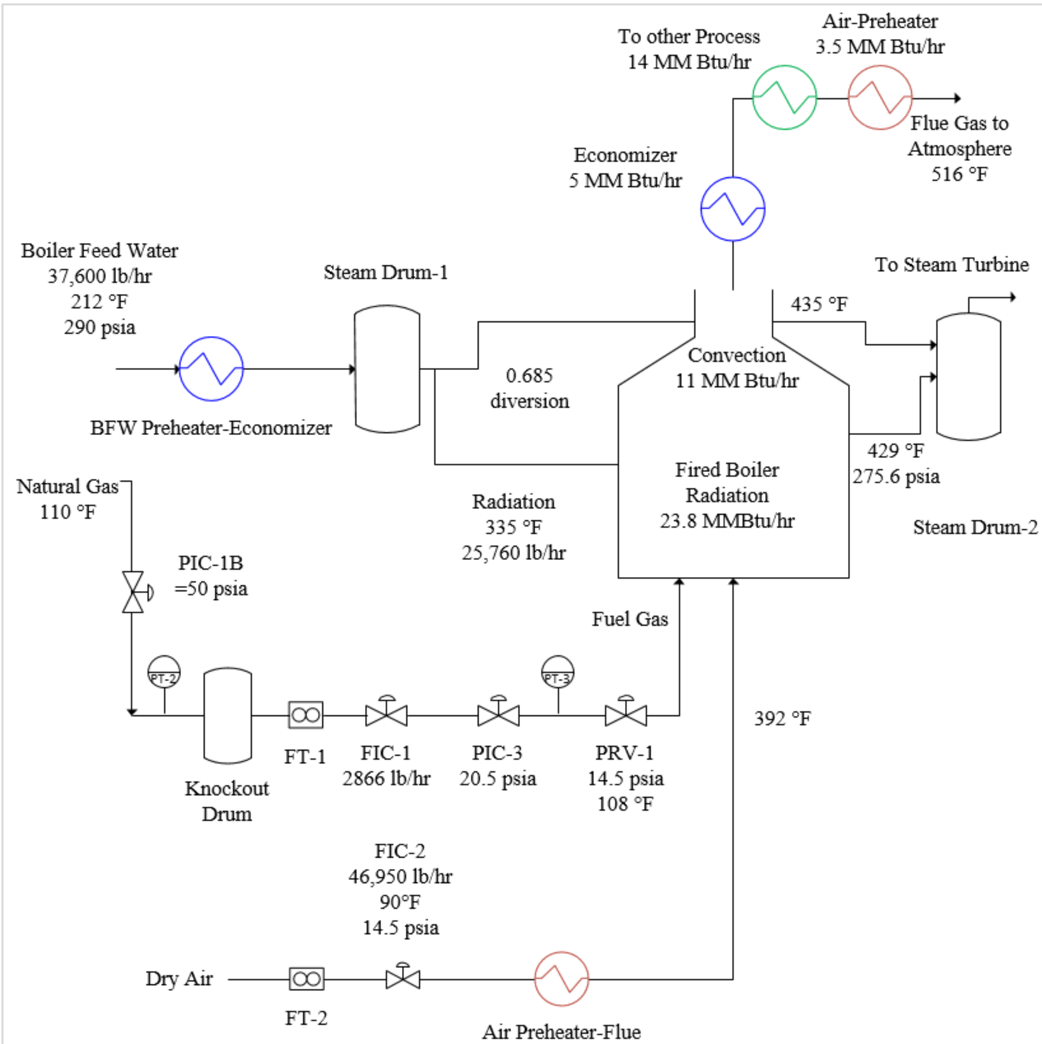


Figure 15 Process Flow Diagram with Base Case

Table 6 Initial Condition of Boiler with Base Case (Natural Gas)

Unit	Natural Gas	Air	BFW	Steam	Boiler Duty (MM Btu/hr.)	External Process (MM Btu/hr.)
Temperature (°F)	110	90	212	429	43.5	14
Pressure (psia)	50	14.5	290	261		
Flow (lb./hr.)	2,866	46,950	37,600	37,600		
Stream to Boiler-R Inlet Temp (°F)	108	392	335			
Stream to Boiler-R Pressure (psia)	14.5	13.8	275.6			

Flare streams from an Ethylene plant is identified as a potential source of high heating value fuel to cogeneration unit. The three flaring sources from Ethylene plant are identified as flare-A, flare B and flare-C with known composition and a flaring frequency of 12 hrs. annually. After integration of flare gas to fire boiler fuel gas system, the pressure of fuel gas system is maintained at 50 psia with sequential control, i.e. primary pressure control is through flare gas system. In case, there is no flaring, fuel gas header pressure is maintained through natural gas.

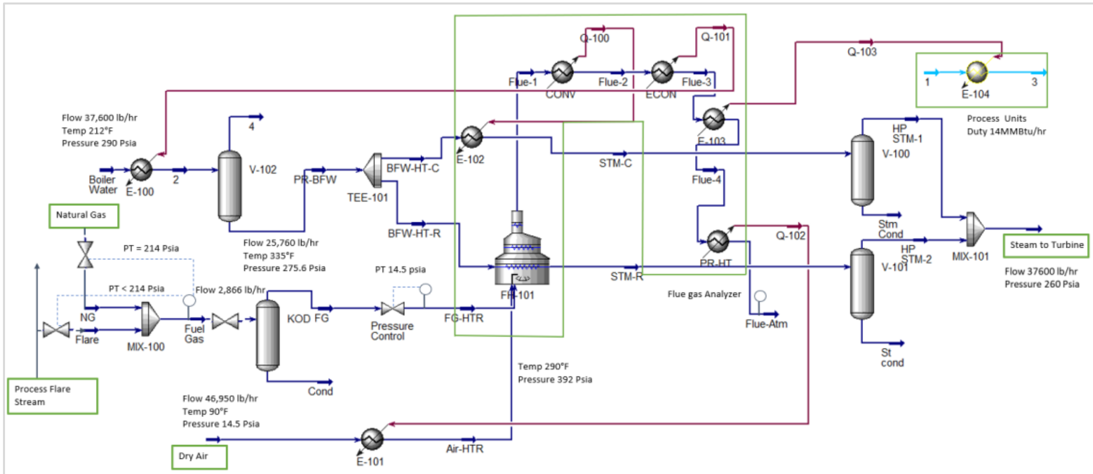


Figure 16 Boiler Simulation Flowsheet after Flare Integration

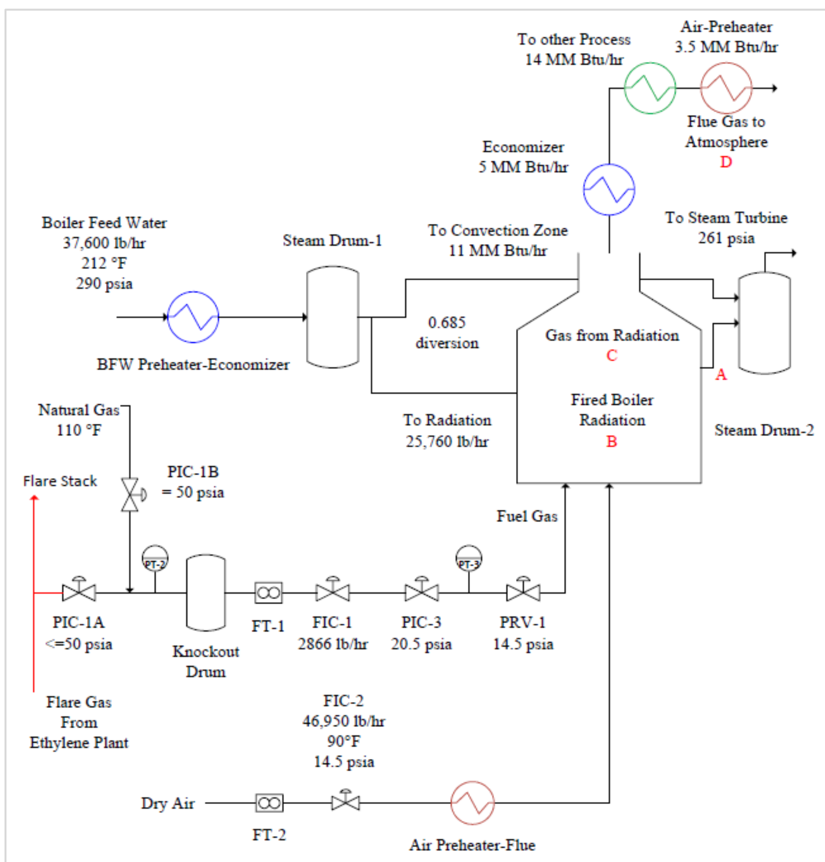


Figure 17 Process Flow Diagram after Integration with Flare Streams

4.2. Process Simulation and Sensitivity Analysis

After the initial simulation of the base case boiler process is completed (shown in Figure 17), the flare stream data is gathered as shown in Table 7.

Table 7 Flare Stream Data (Serveh Kamrava K. J.-H., 2015)

Parameter	Natural Gas	Flare-A	Flare-B	Flare-C
Temperature °F	110	-57	179	-19.40
Pressure psia	50	335	464	270
Mass Flow tons/year	1.03e10	2248	2248	1350
Wobbe Index Btu/SCF	1367	1325	1327	1615
Composition				
Hydrogen	0	0.427	0.423	0
Methane	0.93	0.091	0.092	0
Ethane	0.04	0.069	0.0695	0
Ethylene	0	0.406	0.416	1
C4/C5/N2	0.03	0	0	0

The flare stream data has five parameters – temperature, pressure, flow, heat of combustion and composition. Since, the process has fuel pressure and flow controllers the pressure and flow parameters will not have impact on the system with the defined process control scheme. Heat of combustion is a function of composition. So, temperature and composition will be considered for sensitivity analysis.

The sensitivity analysis and comparison with base case will be carried out based on these five categories:

I. Check with flare stream point values- composition and temperature.

II. Vary fuel gas temperature.

III. Different composition effect on heating value and Wobbe Index

IV. Vary fuel gas composition with constant air flow.

V. Pure composition at stoichiometric air to fuel ratio.

4.2.1. Effect of Flare Stream Point Values on Boiler Performance

The flare stream and natural gas composition data are shown in Figure 18 and Table 7. Natural gas, flare-A, flare-B and flare-C primarily contains hydrogen, methane and ethylene.

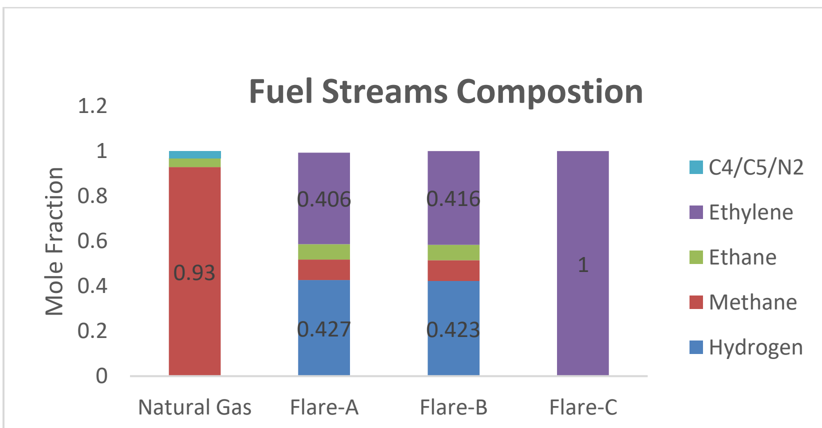


Figure 18 Flare and Natural Gas Composition

Initially, the properties of each major composition (i.e. hydrogen, methane, ethylene, and ethane) is tabulated in Table 8 to later understand their effect on the process.

Table 8 Properties of Fuel Composition

Properties	Hydrogen	Methane	Ethane	Ethylene
Molecular Weight	2.016	16.04	30.07	28.05
Lower Flammability Limit (vol% in air)	4	5.3	3	2.75

Table 8 Continued

Properties	Hydrogen	Methane	Ethane	Ethylene
Upper Flammability Limit (vol% in air)	75	15	12.4	28.6
Minimum Ignition Energy (mJ)	0.02	0.29	0.24	0.07
Higher Heating Value (Btu/lb at NTP)	60,340	23,710	22,180	21,540
Higher Heating Value (Btu/SCF)	310	951	1,717	1,552
Specific Heat (Btu/lb-mole-F at NTP)	6.78	8.57	12.55	10.46
Adiabatic Flame Temp (Mix at 25°F) °F	3860	3547	3608	3806
Stoichiometric Air-Fuel Ratio (mole basis)	2.4	9.7	14.4	15.44
Flame Speed at Stoichiometric (ft/sec at 25°F and equivalence ratio=1)	7.8	1.4	1.46	2.23
Wobbe Index (Btu/ SCF)	1296	1432	1828	1715

Initially, to check the effect of flare point values manipulated variables, observed variables and fixed variables are identified as shown in Table 9. The flow and pressure values of flare gas streams are not incorporated in base case condition because the system already has pressure and flow controller to maintain constant fuel flow and pressure.

After simulation of given fuel stream point values on base case model, rise in boiler radiation zone duty, steam outlet from radiation zone, flue gas temperature from radiation zone and stack gas temperature were identified for flare-A and flare-B containing hydrogen and ethylene in majority.

Table 9 Manipulated, Observed and Fixed Variables for Simulation of Flare Gas Point Values

<table>
<tr><th>Sl. No.</th><th>Manipulated Variables</th><th>Observed Variables</th><th>Fixed Variables</th></tr>
<tr><td>1.</td><td rowspan="4">Point values of flare-A, B, C composition, temperature</td><td>Boiler radiation zone duty</td><td rowspan="4">• Boiler feed water and air flow rate, temperature, pressure.
• Fuel gas pressure, gas mass flow.
• Heat duty of convection, economizer, preheater and to external process.</td></tr>
<tr><td>2.</td><td>Boiler radiation zone outlet steam temperature</td></tr>
<tr><td>3.</td><td>Flue gas to atmosphere temperature</td></tr>
<tr><td>4.</td><td>Boiler radiation outlet flue gas temperature</td></tr>
</table>

Table 10 Simulation Result of Changing Fuel from Natural Gas to Flare Gas

Parameter	Natural Gas	Flare-A	Flare-B	Flare-C
Steam Outlet Temperature (Radiation zone) °F	429	546.8	543	410.4
Flue Gas to Atmosphere °F	516	685.2	681.4	464.2
Radiation Zone Duty MM Btu/hr	23.80	25.3	25.30	23.2
Radiation Outlet Gas °F	2626	2793	2790	2673

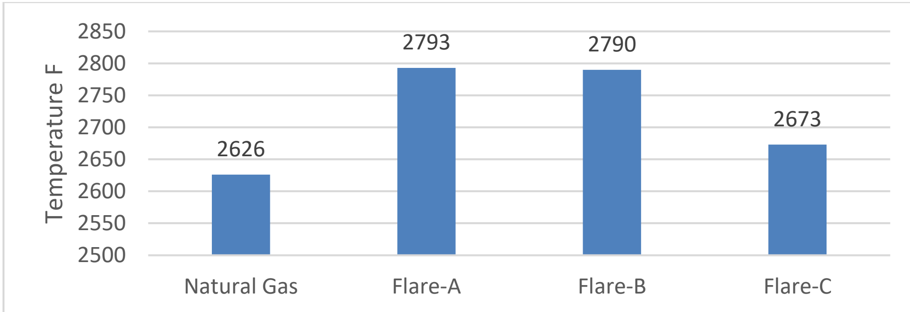


Figure 19 Effect of Flare Gas on Radiation Outlet Gas Temperature

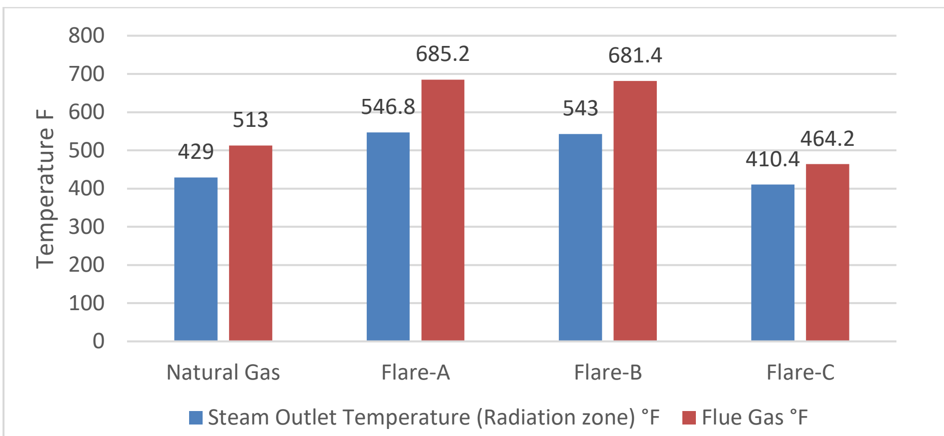


Figure 20 Effect of Flare Gas on Radiation Steam Temperature and Flue Gas Temperature

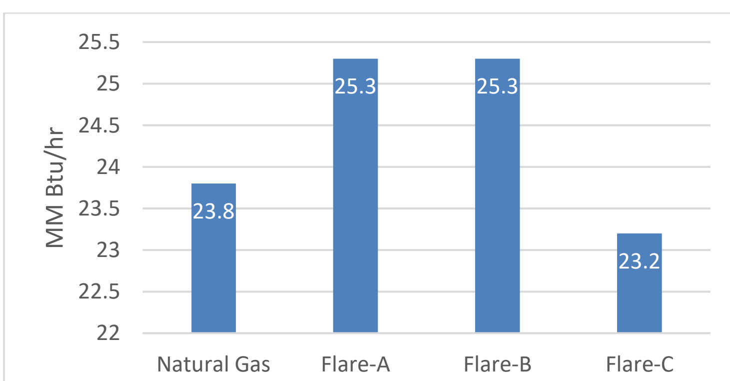


Figure 21 Flare Gas Effect on Boiler Radiation Zone Duty

From the above simulation (shown in Table 10, Figure 19, Figure 20 and Figure 21), it is observed that change in flare gas composition and temperature has an effect on boiler operating conditions. Thus, the study requires further evaluation to identify most influencing parameters effecting the operating condition and their consequences to the cogeneration system.

4.2.2. Effect of Fuel Gas Temperature on Boiler Operation

To check the effect of flare stream temperature on boiler base case, manipulated variables, observed variables and fixed variables are identified as shown in Table 11 .

Table 11 Manipulated, Observed and Fixed Variables for Simulation of Flare Gas Temperature

Sl. No.	Manipulated Variables	Observed Variables	Fixed Variables
1.	Individual temperatures of flare-A, B, C.	Fuel flow of boiler	• Boiler feed water and air flow rate, temperature, pressure. • Fuel gas pressure, gas mass flow. • Heat duty of convection, economizer, preheater and to external process.
2.		Boiler radiation zone outlet steam temperature	
3.		Boiler radiation outlet flue gas temperature	
4.		Boiler radiation zone duty.	

The flare stream temperature is individually simulated with the given flare gas temperature values while the rest of all the fuel gas (including composition), air and boiler water parameters are kept consistent with the base condition. The results are shown in Table 12.

Table 12 Effect of Flare Stream Temperature on Boiler

Parameter	Natural Gas	Flare-A	Flare-B	Flare-C
Fuel to Heater Temperature °F	108	-57.01	179.33	19.39
Fuel Flow to Heater lb/hr	2866	2810	2866	2866
Radiation Outlet Steam Temperature °F	429	410	429	429
Stack Flue Gas to Atmosphere °F	515	460	523	502
Radiation Zone Duty MM Btu/hr	23.8	23.3	23.8	23.8
Radiation Outlet Gas Temperature °F	2626	2587	2633	2616

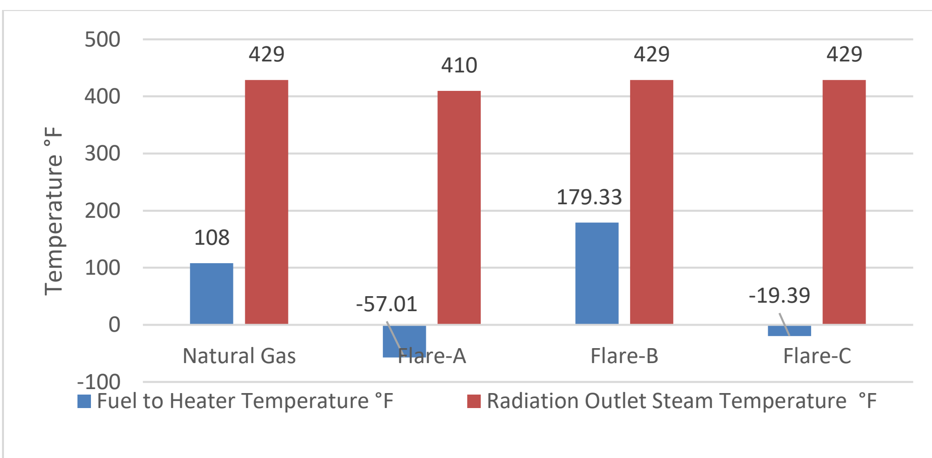


Figure 22 Effect of Fuel Gas Temperature on Radiation Zone Steam Temperature

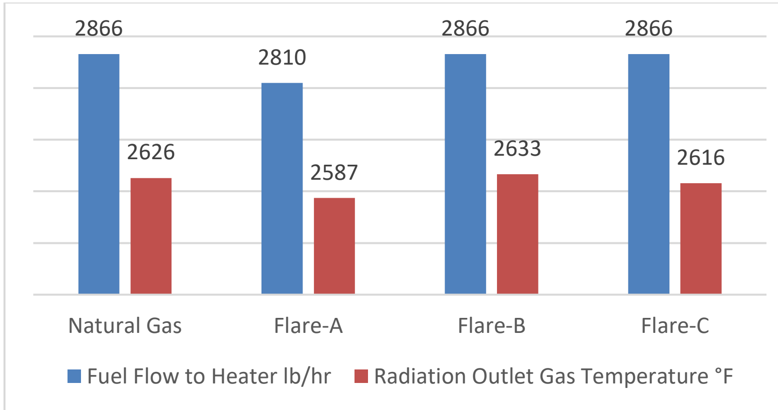


Figure 23 Effect of Fuel Gas Temperature on Fuel Flow and Radiation Gas Temperature

From simulation of fuel temperature, it is observed that if the flare gas temperature is above condensation temperature, there is marginal change in the boiler performance. When the fuel gas header pressure is at 50 psia, if the natural gas temperature drops below -35 °F, part of gas will start condensing and there will loss of condensate with a steady opening of fuel control valve opening (shown in Figure 22 and Figure 23).

4.2.3. Wobbe Index of Different Fuel Composition

To understand the effect of composition on variable composition on boiler performance, it's imperative to check the heating value of variable composition. Hydrogen to carbon ratio and heating value is generally used to characterize a particular fuel composition. For interchangeable fuels, to compare the combustion energy, Wobbe Index is used. It measures the heat input to an equipment through a given aperture at a defined pressure. This means, gas mixtures that have identical Wobbe Index, will deliver the same amount of heat (Speight, 2019).

$$WI = \frac{HHV}{\sqrt{SG}}......\text{Equation 7}$$

For a fixed mass flow of fuel, higher heating value (HHV) is required. The Wobbe Index and HHV is observed at standard temperature (77 °F) and pressure (14.7 psia) for different composition to understand the flame stability and burning profile. Changes in WI indicates flame instability and possibility of greater emissions (Bryan Li, 2009). There are different ranges for Wobbe Index depending on equipment size and power requirement. Gas Safety Management regulations advices a Wobbe Index range 1248 to 1418 BTU/SCF at 60°F in United Kingdom (Haywood, 2011).

The manipulated variables, observed and fixed variables are shown in Table 13.

Table 13 Manipulated, Observed and Fixed Variables for Composition Heating Value and Wobbe Indices

Sl. No.	Manipulated Variables	Observed Variables	Fixed Variables
1.	Composition- Hydrogen, Methane and Ethylene.	Heating Value of gas.	• Boiler feed water and air flow rate, temperature, pressure. • Fuel gas temperature, pressure, mass flow. • Heat duty of convection, economizer, preheater and to external process.
2.		Wobbe Indices of gas.	

Table 14 Higher Heating Value and Wobbe Index for Hydrogen and Methane Mixture

Hydrogen	1	0.9	0.85	0.75	0.7	0.6	0.45	0.35	0.2	0.05	0
Methane	0	0.1	0.15	0.25	0.3	0.4	0.55	0.65	0.8	0.95	1
HHV Btu/lb	60,330	43,150	38,940	33,740	32,000	29,500	27,000	26,000	24,800	23,950	23,700
Wobbe Index BTU/SCF	1,174	592	534	463	1,095	1,118	1,155	1,181	1,238	1,284	1,303

Table 15 Higher Heating Value and Wobbe Index for Methane and Ethylene Mixture

CH_4	0.9	0.8	0.6	0.5	0.4	0.3	0.2	0
C_2H_2	0.1	0.2	0.4	0.5	0.6	0.7	0.8	1
HHV Btu/lb	23,400	23,000	22,540	22,300	22,100	21,970	21,800	21,500
Wobbe Index BTU/SCF	1,334	1,364	1,410	1,439	1,465	1,493	1,519	1,566

Table 16 Higher Heating Value and Wobbe Index of Ethylene and Hydrogen Mixture

H2	0	0.1	0.25	0.4	0.5	0.6	0.75	0.8	0.85	0.9	1
C2H2	1	0.9	0.75	0.6	0.5	0.4	0.25	0.2	0.15	0.1	0
HHV Btu/lb	21500	21850	22450	23300	24140	25320	28420	30200	32770	36780	60330
Wobbe Index BTU/SCF	1566	1516	1432	1343	1285	1225	1138	1113	1094	1084	1174

Table 14, Table 15, Table 16, Figure 24 and Figure 25 shows the flame stability and heating capacities of major composition in flare gas. Higher heating value is observed for higher mole fraction of hydrogen due to lower density (higher volumetric flow at constant mass flow rate).

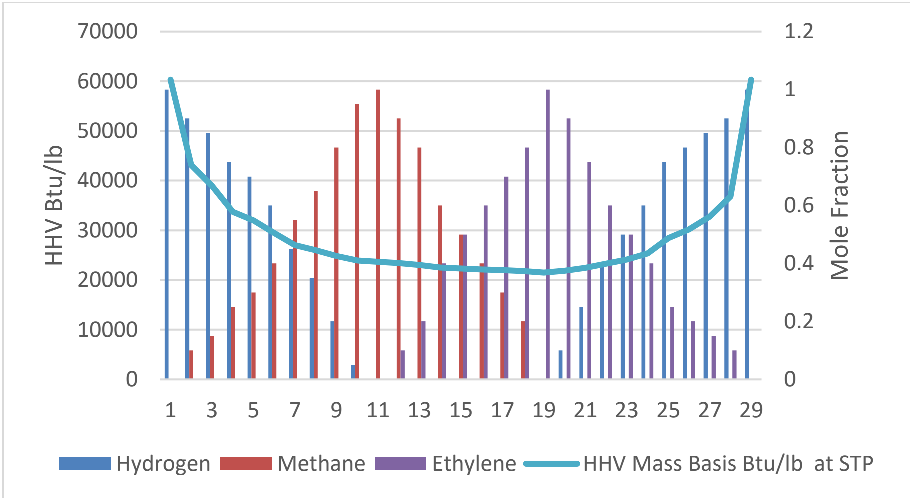


Figure 24 Composition Effect on Heating Value Mass Basis

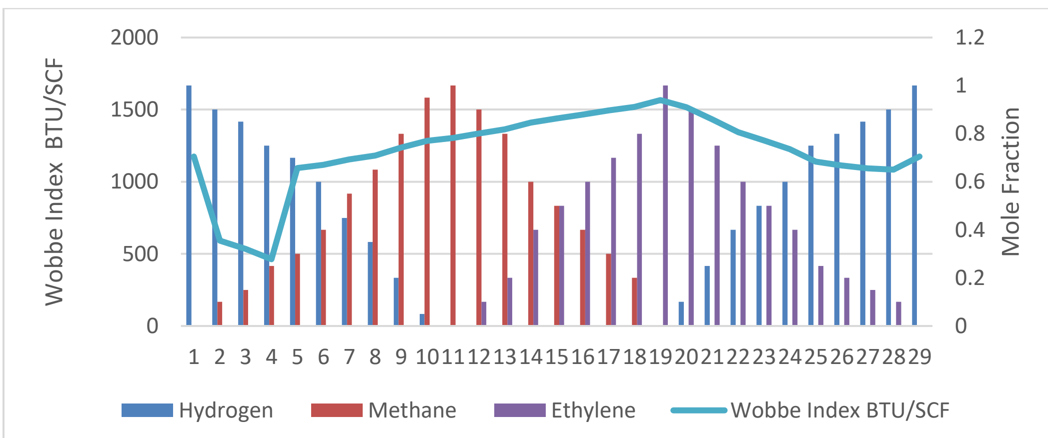


Figure 25 Wobbe Index for Different Composition

In base case, the heating value was 22,900 Btu/lb at STP and Wobbe Index was 1316 Btu/SCF. If the fuel flow controller is stuck at same opening (rendering same volumetric flow for all compositions), then with hydrogen above 20 mole% with methane and 60 mole% with ethylene, Wobbe Index is lower than 1250 Btu/SCF. This situation will lead to flame instability as observed at higher H2 concentration (above 60 mole% H2 to 90 mole% H2).

4.2.4. Effect of Fuel Composition on Boiler Operation

The data collected shows flare streams primarily contains hydrogen, methane and ethylene. Thus, these three compositions are simultaneously changed, while the fuel temperature, pressure, flow as well as air and boiler water conditions are kept constant. The manipulated variables, observed and fixed variables are shown in Table 17. The results are shown in Table 18, Table 19, and Table 20.

Table 17 Manipulated and Observed Variable for Fuel Composition Change Analysis

Sl. No.	Manipulated Variables	Observed Variables	Fixed Variables
1.	Composition of fuel stream – Hydrogen, Methane and Ethylene	Boiler radiation zone outlet steam temperature	• Boiler feed water and air flow rate, temperature, pressure. • Fuel gas temperature, pressure, mass flow. • Heat duty of convection, economizer, preheater and to external process.
2.		Boiler radiation zone duty	
3.		Stack flue gas to atmosphere temperature	
4.		Composition of stack flue gas	

Table 18 Hydrogen and Methane Composition Effect Analysis

Hydrogen Gas and Methane Composition v/s Boiler Performance								
Hydrogen	1	0.85	0.75	0.6	0.45	0.3	0.15	0
Methane	0	0.15	0.25	0.4	0.55	0.7	0.85	1
Steam-Radiation Outlet °F	809.0	721.2	535.0	410.0	410.0	410.0	410.4	410.4
Furnace Flue Gas Outlet °F	2240	2597	2523	2473	2516	2544	2564	2579
Flue Gas to Atmosphere °F	742.8	801.0	601.4	449.0	457.4	462.8	466.6	469.5
Boiler-Radiation Duty MMBTU	28.80	27.60	25.19	23.40	23.40	23.40	23.40	23.40
CO in Flue Gas	-	0.016	0.05	0.07	0.048	0.033	0.022	0.014

Table 19 Ethylene and Methane Composition Change Effect Analysis

Ethylene Gas and Methane Composition v/s Boiler Performance									
Methane	1.00	0.95	0.85	0.75	0.60	0.45	0.30	0.15	0.00
Ethylene	0.00	0.05	0.15	0.25	0.40	0.55	0.70	0.85	1.00
Steam-Radiation Outlet °F	410.4	421.2	462.4	452.4	434.5	419.4	410.4	410.4	410.4
Furnace Flue Gas Outlet °F	2579	2613	2675	2679	2677	2676	2675	2674	2674

Table 19 Continued

Ethylene Gas and Methane Composition v/s Boiler Performance									
Flue Gas to Atmosphere °F	469.5	502.6	562.6	550.7	528.2	508.9	492.0	477.5	464.5
Boiler-Radiation Duty MMBTU	23.40	23.70	24.24	24.11	23.87	23.68	23.50	23.36	23.24
CO in Flue Gas	0.014	0.009	0.0006	-	-	-	-	-	-

Table 20 Ethylene and Hydrogen Composition Change Effect Analysis

Ethylene and Hydrogen Composition v/s Boiler Performance											
Hydrogen	0	0.1	0.25	0.4	0.5	0.6	0.75	0.8	0.85	0.9	1
Ethylene	1	0.9	0.75	0.6	0.5	0.4	0.25	0.2	0.15	0.1	0
Steam-Radiation Outlet °F	410	410	450	512	572	647	628	616	606	636	809
Furnace Flue Gas Outlet	2673	2693	2727	2776	2822	2876	2782	2730	2662	2585	2240
Flue Gas to Atmosphere °F	463	497	555	637	713	808	758	731	701	709	742
Boiler Duty-Radiation MM Btu/Hr	23.2	23.5	24.1	24.9	25.7	26.7	26.4	26.3	26.1	26.5	28.8
CO In Flue	-	-	-	-	-	0.001	0.035	0.054	0.073	0.060	-

I. Comparison of boiler outlet steam with composition fuel gas stream.

In base case, radiation steam temperature was 429 °F. Figure 26 shows radiation steam outlet temperature rising above 500°F when hydrogen is higher than 75 mole% in the methane- hydrogen mixture and 40 mole% in ethylene-hydrogen mixture. The rise in temperature in radiation section are contributed by higher molar flow of hydrogen at constant mass flowrate. Moreover, hydrogen has high flame speed and high heating value (mass basis 60,340 Btu/lb). If steam consumption to turbine is constant, and the water siphon tubes to radiation section are maintained at elevated temperature, excessive pressure in the steam drum can lead to steam explosion. When methane is 80 mole% and ethylene at 20 mole%, rise in temperature is due the combustion mixture reaching stoichiometric air-fuel ratio.

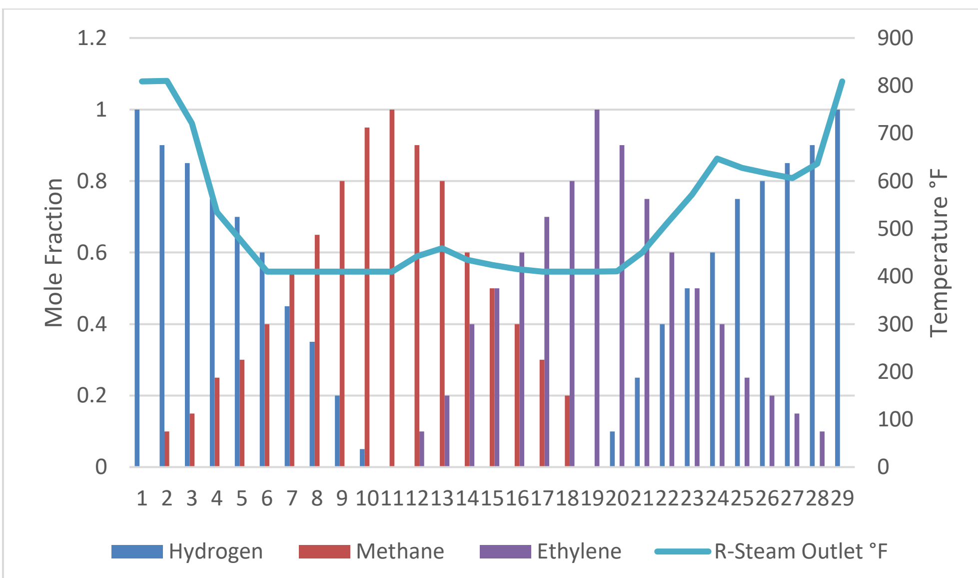


Figure 26 Fuel Composition Effect on Boiler Steam Outlet Temperature

II. Composition Variation Effect on Stack Flue Gas Temperature

In base case, stack flue gas temperature was at 515 °F. The temperature of flue gas follows the same trend on variation of composition as radiation outlet steam temperature. At pure hydrogen, the gas mixture in the firebox is too rich. At 75 mole% hydrogen with methane and 35 mole% hydrogen with ethylene, temperature rise above 600 °F is observed due to the presence of excess hydrogen (refer to Figure 27).

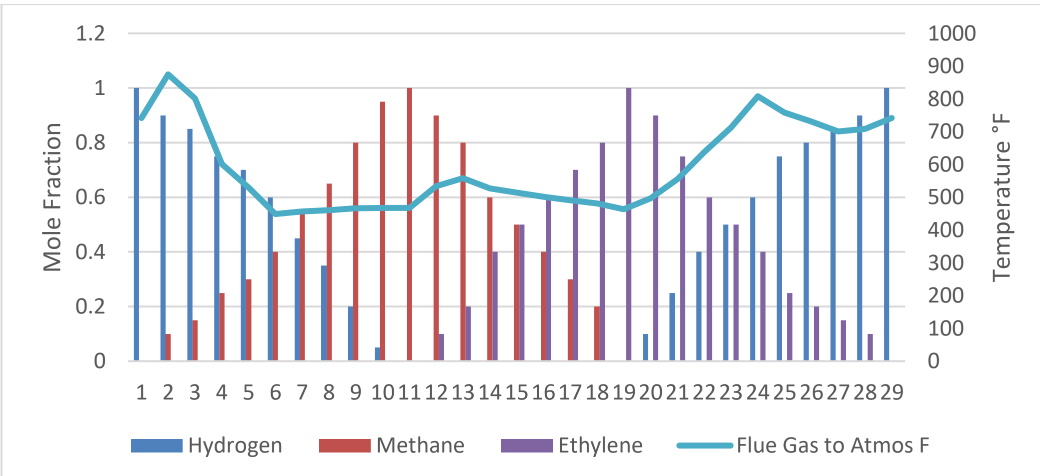


Figure 27 Fuel Composition Effect on Stack Flue Gas temperature

III. Composition Variation Effect on Boiler Radiation Zone Duty

In base case, boiler duty was 23.8 MM Btu/hr. As shown in Figure 28, the boiler zone duty has increased above 24.8 MM Btu/hr. when pure hydrogen was introduced at higher molar rate (more than 75 mole% hydrogen with methane and 40 mole% with ethylene).

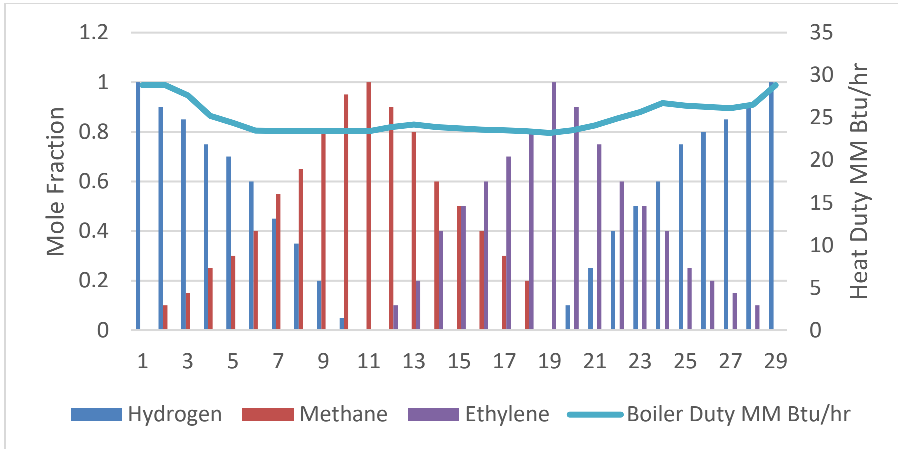


Figure 28 Fuel Composition Effect on Boiler Radiation Duty

IV. Composition of flue gas on varying boiler fuel gas stream.

In base case, flue gas stack had minute CO content (0.0008). However, Figure 29 shows presence of methane in flare stack gas, when hydrogen composition (rest methane) in fuel gas is above 65 mole%. At elevated flare stack temperature, unburnt hydrocarbons can ignite and can finally lead to flare stack explosion.

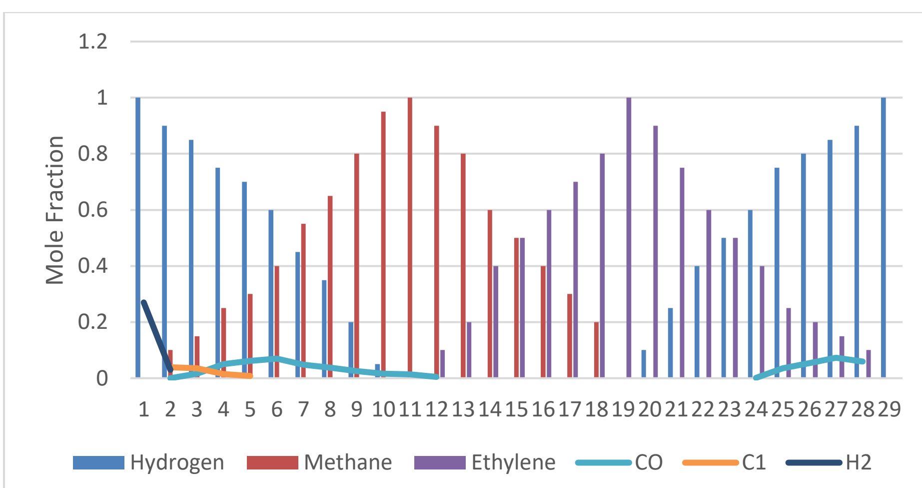


Figure 29 Fuel Composition Effect on Flue Gas Composition

In base case with air-fuel ratio of 9.92, there was CO of 0.0008. On varying composition, rich fuel condition is observed at higher hydrogen molar flow and as we introduce higher molecular weight hydrocarbons, the molar rate reduces and the fuel becomes leaner. Carbon monoxide and oxygen curve is shown in Figure 30, which indicate when methane is 80 mole% with ethylene and hydrogen at 55 mole% with ethylene, the combustion mixture is in stoichiometric condition, which minimal fuel consumption and optimum energy output.

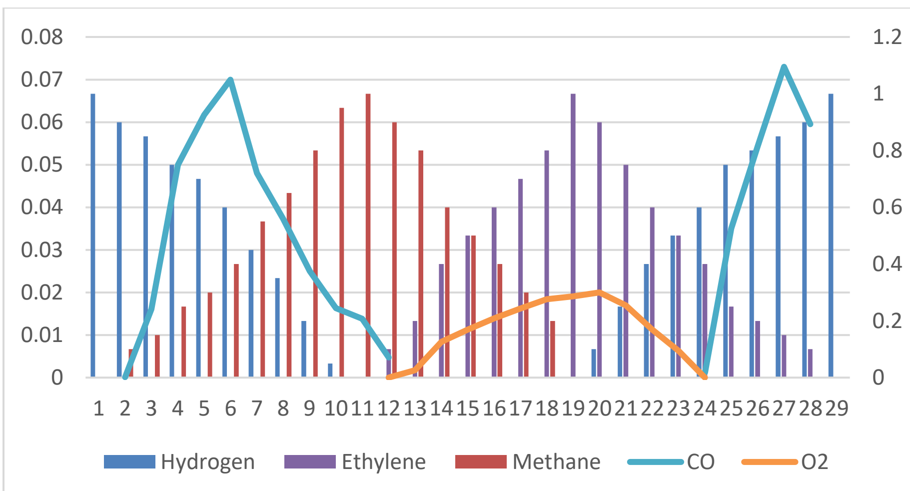


Figure 30 Carbon Monoxide and Oxygen in Flue Gas

4.2.5. Change of Air with Change in Fuel Gas Composition

Air-fuel ratio is a common term used to describe the mixing of air and fuel in the combustion zone. Stoichiometric air to fuel ratio (mole basis) for hydrogen is 2.39, methane is 9.52 and ethane is 16.68.

$$AFR = \frac{\boldsymbol{Molar\ mass\ (or\ volume)\ of\ air}}{\boldsymbol{molar\ mass\ (or\ volume)\ of\ fuel\ gas}} \text{ … …Equation 8}$$

Air is increased to stoichiometric value based on the pure composition (hydrogen, methane, ethylene) to identify the changes in the process. This is to investigate, if pure compositions are in the fuel gas system and air is increased accordingly, how it is affecting the process. Also, to compare with base case and identify lean and rich fuel mixture formed. The manipulated variables, observed and fixed variables are shown in Table 21 and the simulation results in Table 22 and Table 23.

Table 21 Manipulated Observed and Fixed Variables for Stoichiometric Air to Fuel Ratio for Pure Components

<table>
<tr><th>Sl. No.</th><th>Manipulated Variables</th><th>Observed Variables</th><th>Fixed Variables</th></tr>
<tr><td>5.</td><td rowspan="4">Air and Pure Composition of Hydrogen, Methane and Ethylene</td><td>Boiler radiation zone outlet steam temperature</td><td rowspan="4">Boiler feed water and air flow rate, temperature, pressure.
Fuel gas temperature, pressure, mass flow.
Heat duty of convection, economizer, preheater and to external process.</td></tr>
<tr><td>6.</td><td>Boiler radiation zone duty</td></tr>
<tr><td>7.</td><td>Stack flue gas to atmosphere temperature</td></tr>
<tr><td>8.</td><td>Boiler radiation zone outlet gas temperature</td></tr>
</table>

Table 22 Pure Composition at Stoichiometric Air-Fuel Ratio

Fuel Mass Flow =2866 lb/hr	**Hydrogen**	**Methane**	**Ethane**	**Ethylene**	**Base Case**
Air Flow lb/hr	9.75E+04	5.00E+04	4.62E+04	4.25E+04	4.71E+04
Air Flow lb-mole/hr	3378	1732	1472	1472	1632
Air Fuel Ratio	2.40	9.70	15.44	14.40	9.95
Radiation Outlet Steam °F	2808	491.3	410.4	410.4	434
Radiation Flue Gas Outlet °F	2733	2570	2656	2877	2626
Heater Duty MMBTU/hr	59.3	24.6	23.38	23.23	23.87
Flue Gas to Atmosphere °F	1854	491.4	478.5	497.9	520
O2 in Flue Gas to Atmos. Mole percent	0.0013	0.0052	0.0033	0.0036	0.0002

Table 23 Pure Composition at Constant Air Flow

100% Composition with Base Conditions	**Hydrogen**	**Methane**	**Ethylene**	**Ethane**
Air Flow lb/hr	4.70E+04	4.70E+04	4.70E+04	4.70E+04

Table 23 Continued

100% Composition with Base Conditions	Hydrogen	Methane	Ethylene	Ethane
Air Flow lb-mole/hr	1627	1627	1627	1627
Air Fuel Ratio	1.10	9.11	15.92	17.07
Radiation Outlet Steam °F	809	410.4	410	410.4
Radiation Flue Gas Outlet °F	2240	2579	2674	2625
Heater Duty MMBTU/hr	28.8	23.4	23.24	23.38
Flue Gas to Atmosphere °F	742.8	469.5	464.5	473.3
O2 in Flue Gas to Atmos. Mole percent	0	0	0.0222	0.0063
CO in Flue Gas to Atmos. Mole percent	0.2707 H2	0.0138	-	-
Type of Combustion	Rich Fuel	Rich Fuel	Lean Fuel	Lean Fuel

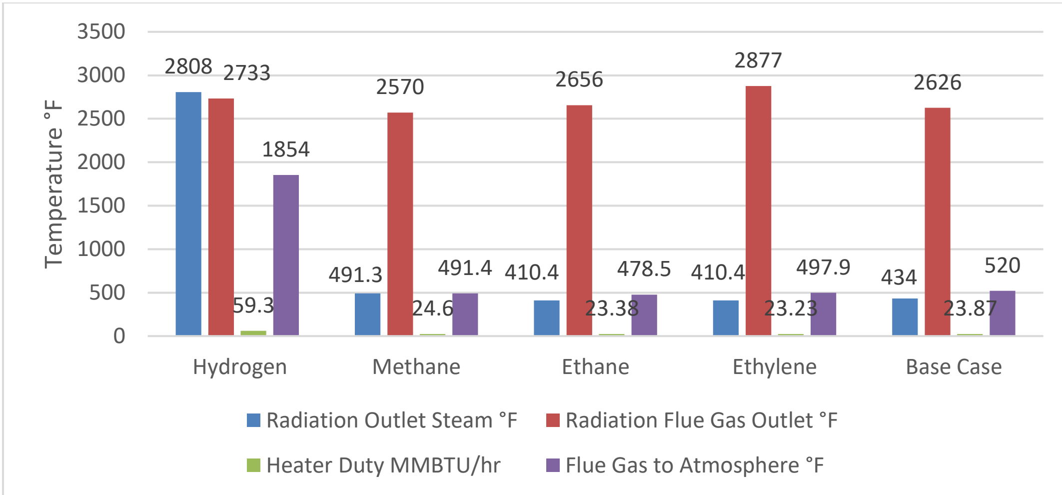


Figure 31 Comparison of Pure Composition at Stoichiometric Air Fuel Ratio with Base Case

Equation 9 and Equation 10 shows radiant section and tube wall temperature (Hassan Al-Haj Ibrahim M. M.-Q., 2013).

$$\boldsymbol{Q = \sigma.(\alpha.A).F.\left(T_g^4 - T_w^4\right)} \ldots\ldots \text{Equation 9}$$

$$\boldsymbol{T_w = 100 + 0.5\left(\frac{T_i + T_o}{2}\right)} \ldots\ldots \text{ Equation 10}$$

Where, σ = Stefan-Boltzman constant (2.041 x 10E-7 kJ/h m^2k^4)

α = Effectiveness factor of tubes bank

A = Heat exchange surface area (m^2)

F = Exchange Factor (0.97)

T_g = Gas temperature in firebox (also flame temperature) (K)

T_w = Average tube wall temperature (K)

T_i = Process fluid inlet temperature (K)

T_o = Process fluid outlet temperature (K)

The radiation duty is more than double if pure hydrogen is present. Figure 31 shows the elevated steam temperature from radiation zone and radiation zone duty. From equation 9 and Equation 10, it is shown that when the radiation zone temperature and radiation zone heat duty is high, the tube wall temperature will be high, which can lead to tube wall rupture, flame impingement. When pure ethylene is present, the steam temperature is reaching the steam saturation temperature. While transporting to turbines, loss of energy can lead to condensate formation in pipelines.

4.2.6. Analysis and Hazard Identification

From simulation and sensitivity analysis of flare composition and temperature, following inferences can be made:

- Identified flare gas parameters can affect the boiler operating conditions. While simulating the flare gas composition and temperature, the boiler radiation zone heat duty and process steam and flue gas temperature changes.
- Flare gas temperature variation has minimal effect on boiler process conditions. When the temperature of fuel gas is below -35 °F, condensate starts to form, which will be separated from the Knockout Drum and the flow can be compensated with fuel control valve opening.
- With variable composition of flare gas, the heating value and Wobbe Index changes. With a constant mass flow, the heating value mass basis is higher for hydrogen (60330 Btu/lb), producing more heat in the firebox area. For ethylene,

lower heat transfer is expected. However, if the fuel flow controller is stuck at same opening (rendering same volumetric flow for all compositions), then with hydrogen above 20 mole% with methane and 60 mole% with ethylene, Wobbe Index is lower than 1250 Btu/SCF. This situation will lead to flame instability as observed at higher H2 concentration (above 60 mole% H2 to 90 mole% H2).

- Pure hydrogen to boiler spiked up the radiation zone steam temperature by 400 °F and flue gas temperature by 270°F (compared to pure methane fuel). Above 40 mole% hydrogen with ethylene the radiation steam temperature started increasing above 500 °F compared to 430 °F in base case condition. Sharp rise in radiation zone steam is observed when hydrogen above 75 mole % is introduced with methane. The rise of sudden process stream (steam) temperature and bridge-wall temperature (flue gas leaving radiant section) indicates high heat transfer in radiation zone. Rise in heat radiant heat transfer raises the fire box flame temperature and rise in process steam temperature increases the tube wall temperature. which can affect the fire box temperature, flame impingement to boiler tubes and radiant section tube metal temperature (or skin temperature). The sudden rise of boiler temperature can also affect the pressure near the arch or the radiation to convection section. The operational draft near the arch is the lowest and any rise can lead to flue gas losses to atmosphere or can cause back fire (if damper is not operated to control boiler draft) as the flow of air will be reduced. The sensitivity analysis of boiler pressure could not be carried out on arch section, but rise in boiler temperature is an indirect indication of its effect on boiler pressure condition.

- On introduction of higher composition of hydrogen (more than 65 mole% with methane), hydrogen and methane carryover were observed in the flue gas stream, due to incomplete combustion. Moreover, higher stack flue gas temperature (more than 600 °F) was observed at higher hydrogen concentration, rendering loss of heat as well as chances of stack explosion in presence of an ignition source.
- The saturation steam temperature of steam at 261 psig is 410 °F. When the fuel gas stream is lighter (methane is above 40% with hydrogen) and when fuel stream is heavy (ethylene more 70 mole% with methane), radiation zone steam temperature is near the borderline of saturation temperature (410.4 °F). If the frictional losses are considered in the pipelines, there is high possibility of condensate carryover to steam turbine.
- During pure hydrogen, air-fuel ratio (1.14) was below the stoichiometric ratio (2.39). Fuel rich firebox can lead to explosion if fresh air is increased too swiftly to compensate for lower oxygen in firebox, because the incoming excess oxygen will rapidly mix with hot unburned fuel. The explosions due to fuel rich conditions usually observed during quick transition from rich to lean burning. When air is stoichiometrically increased for hydrogen, the radiation zone duty (59 MM Btu/hr) has increased by more than double compared to base case (23.8 MM Btu/hr), indicating very high flame temperature, which can affect the integrity of boiler tubes.

Therefore, from the sensitivity analysis, it was observed that hydrogen composition in fuel gas (from flare) and air to fuel ratio are the major parameters affecting boiler operation. Boiler radiation zone temperature and incomplete combustion are the critical scenarios observed from the variation of composition.

4.3. Scenario Development

For scenario development, fault tree and event tree are used for individual top events. The major critical events identified are – boiler flue gas temperature high and incomplete combustion. The change in fuel stream composition is updated (highlighted in red) in both the fault tree diagram.

4.3.1. Fault Tree

Fault tree diagram is made with boiler temperature exceeding design temperature and rich fuel in firebox as two top events (which are identified as critical operational events from simulation). The base events, leading to intermediate and the top event are added to the fault tree. For example, failure of preventive safeguards (e.g. emergency shutdown activation), human error (e.g. operating a wrong valve), software failures (e.g. controller logic), mechanical integrity (loss of cooling due to pump shutdown). The technical hazards (e.g. excess hydrogen) identified due to change in operation updated to the fault tree. Refer to Figure 32 and Figure 33 for the respective fault tree diagrams.

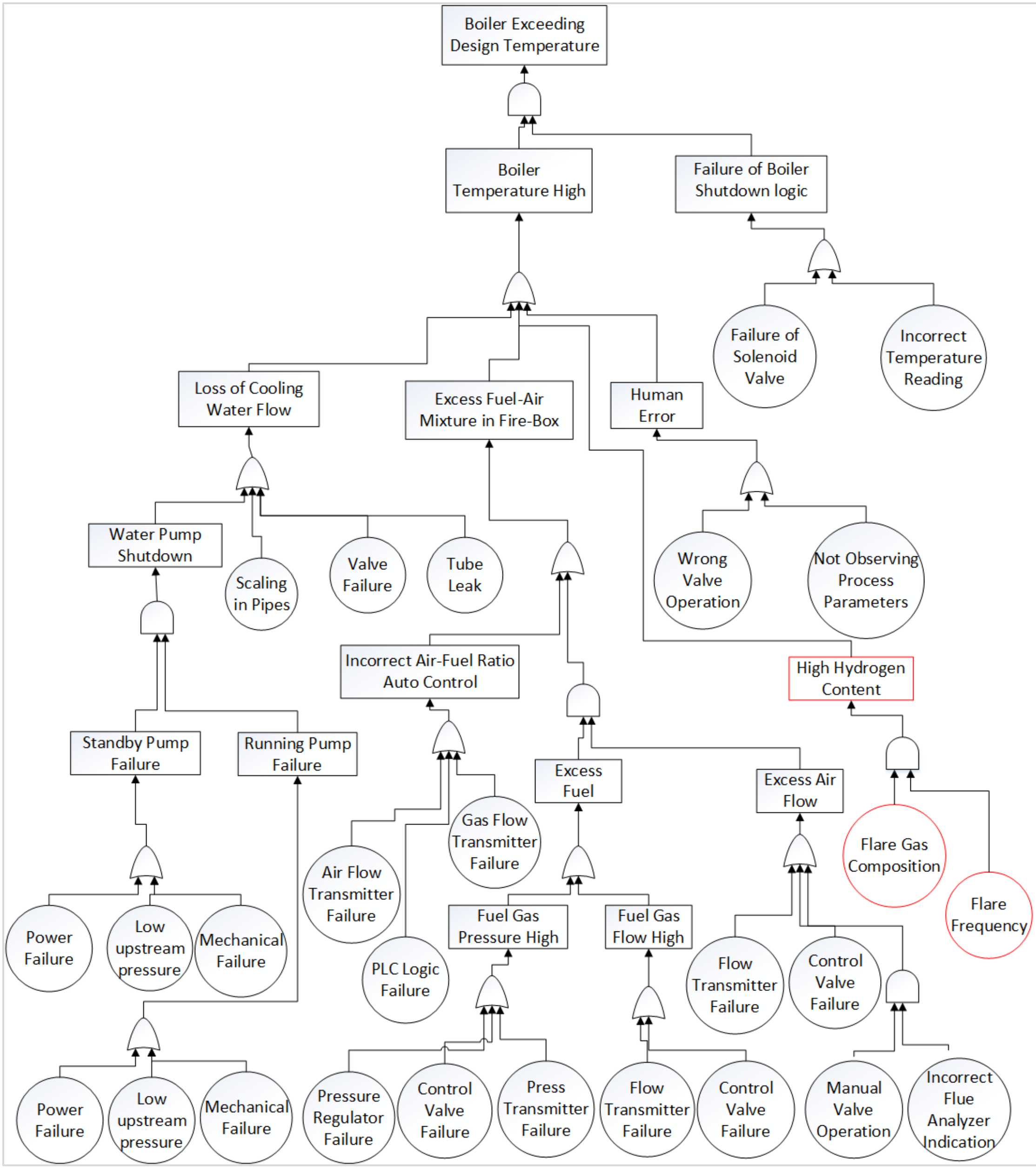


Figure 32 Fault Tree Diagram for Boiler Gas temperature Exceeding Design Temperature

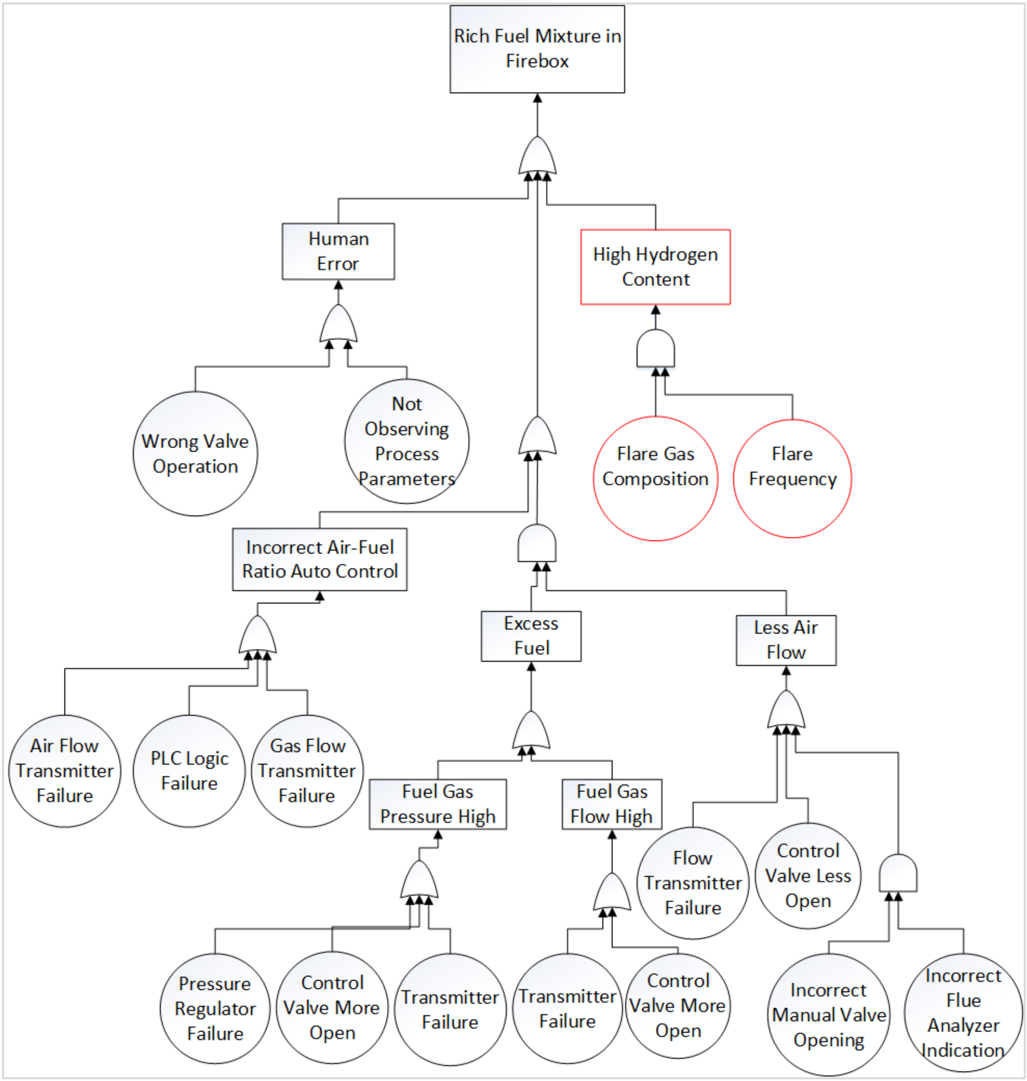


Figure 33 Fault Tree Diagram for Rich Fuel Mixture in Firebox

4.3.1.1. Event Tree

Fault tree is complimented by Event tree. For the event tree, the initiating event is boiler gas temperature exceeding design limit and rich fuel gas mixture in firebox. Subsequently, the safety functions or the mitigating barriers are identified to control the propagation of top event to incident. Figure 34 shows the event tree for boiler gas temperature high. Four barriers were identified – high temperature alarm, manual isolation of fuel valve and steam drum relief vale operation with their corresponding failure probability values collected from OREDA and Lees Loss Prevention in Process Industries database. Four consequences were observed – normal operation, safe shutdown, tube rupture and steam explosion.

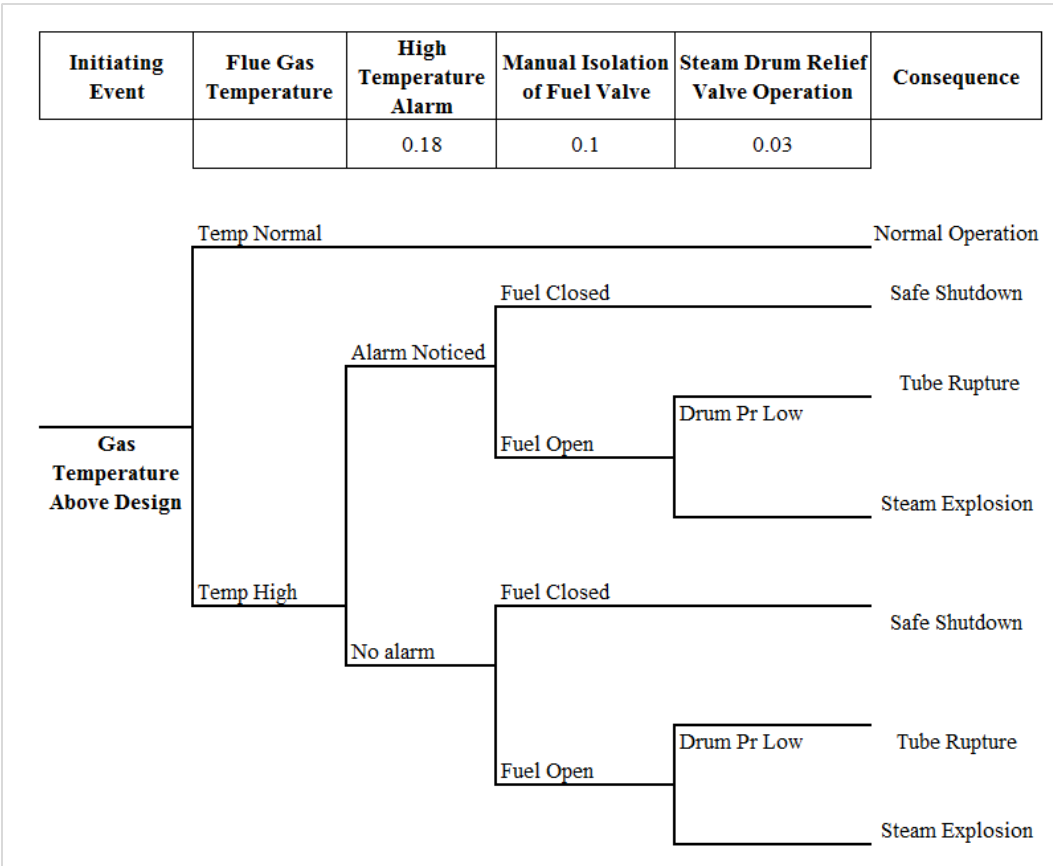


Figure 34 Event Tree for High Boiler Gas Temperature

Figure 35 shows the Event Tree diagram for rich fuel in boiler firebox. Three mitigating safety barriers were identified – stack flue gas composition analyzer, fuel valve operation and air register manual adjustment with their corresponding failure probability (valve and air adjustment probability values are based on operator's failure to adjust the valves in 30 minutes from OREDA database). Three consequences were observed for rich fuel in firebox as initiating event – normal operation, flame impingement, and flameout or stack explosion.

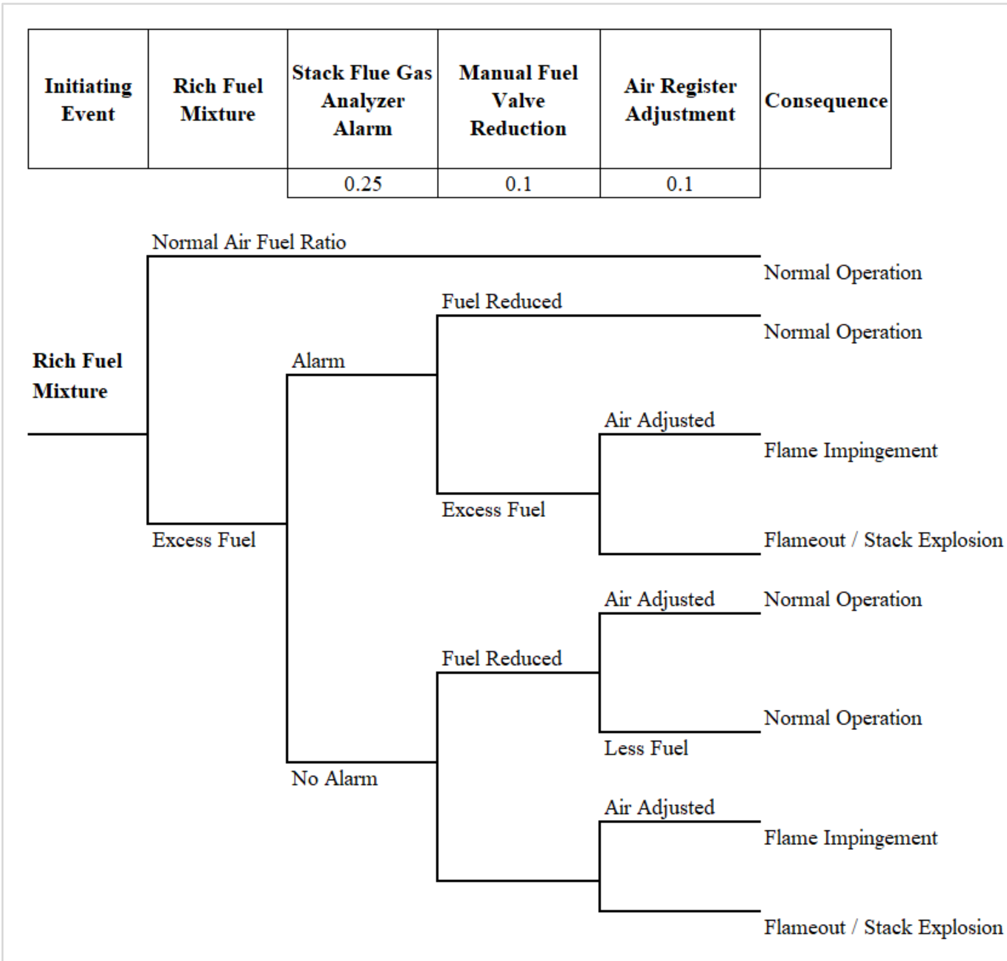


Figure 35 Event Tree for Rich Fuel Incomplete Combustion

4.4. Quantitative Risk Assessment Through Bayesian Network

As discussed in section 3.2, Bayesian Network can be used for diagnosis and prediction analysis. Provided with an evidence of critical event (e.g. boiler flue gas temperature high), the most probable cause can be identified/diagnosed (e.g. flare gas composition). Additionally, the common cause (e.g. power failure effect on standby and running pump) can also be represented in Bayesian Network.

The failure rate data for the components identified in fault tree and event tree are collected from Offshore and Onshore Reliability Data (OREDA) and Lees Loss Prevention in the Process Industries. The data are enumerated in Table 24.

Table 24 Failure Rate of Components

Sl. No.	Node detail	Failure Rate (yr-1)	Failure Probability	Reliability
1	Control Logic for Shutdown	0.250	0.221	0.779
2	Temperature Reading	0.019	0.019	0.981
3	Solenoid Function	0.263	0.231	0.769
4	Operator on Critical Response	0.040	0.039	0.961
5	Wrong Valve Operation	0.050	0.049	0.951
6	Pressure Regulator	0.166	0.153	0.847
7	Pressure Transmitter	0.004	0.004	0.996
8	Pressure Control Valve	0.167	0.153	0.847
9	Flow Transmitter	0.032	0.031	0.969
10	Fuel Gas Flow Control Valve	0.167	0.153	0.847
11	Manual Valve Opening	0.050	0.049	0.951
12	Flue Analyzer Indication	0.289	0.251	0.749
13	Air Flow Control Valve	0.167	0.153	0.847
14	Air Flow Transmitter	0.032	0.031	0.969
15	Process Control Logic for Normal Operation	0.048	0.047	0.953
16	Pump Mechanical Performance	0.042	0.041	0.959
17	Low Pressure Water	0.042	0.041	0.959
18	Power supply with dual UPS (Assumed)	0.00051	0.0005	0.995
19	Pipe Leak Near Pump	0.083	0.080	0.920
20	Water Valves	0.061	0.060	0.940

Table 24 Continued

Sl. No.	Node Detail	Failure Rate (yr-1)	Failure Probability	Reliability
21	Manual Valve Isolation	0.105	0.1	0.900
22	Steam Relief Valve Operation	0.030	0.030	0.970
23	Stack Damper	0.041	0.040	0.960
24	Explosion Door	0.030	0.030	0.970
25	Alarm Indication	0.200	0.18	0.82
26	Gas Analyzer	2.500	0.918	0.082
27	Air Register Opening in 30 minutes	0.105	0.1	0.9

The components are assumed to be operating in homogeneous operating conditions and the plant is operating in useful life phase with constant failure rate. The power reliability of the plant is assumed good and approximated to be 0.005 per year. Flare gas frequency is considered 12 hours/year (from literature). Both the event tree and fault tree are mapped together in Bayesian Network to show the holistic cause-consequence model.

The top event and the consequences identified are shown in Table 25. The whole Bayesian network after mapping fault tree and event tree is shown in Figure 36.

Table 25 Top Event and Consequence in Bayesian Network

Sl. No.	Top Event	Consequence
1.	Boiler Flue Gas Temperature High	Normal Operation
2.		Safe Shutdown
3.		Tube Rupture
4.		Steam Explosion

Table 25 Continued

Sl. No.	Top Event	Consequences
5.	Rich Fuel Mixture in Boiler Combustion Zone	Normal Operation
6.		Flame Impingement
7.		Flameout or Stack Explosion

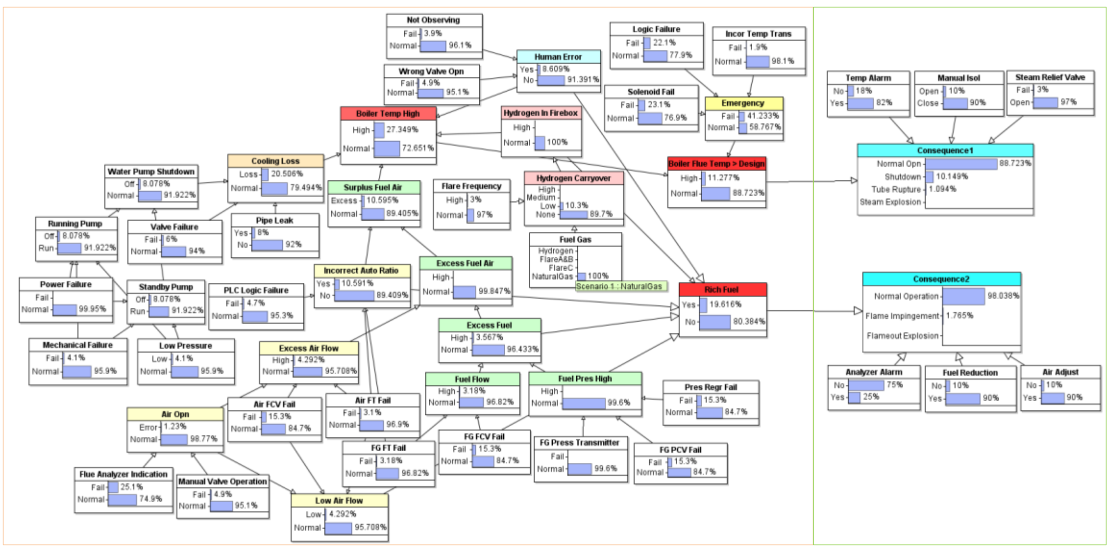


Figure 36 Bayesian Network for Base Case

The frequency of flaring is 12 hrs/year, which is considered as 0.033 probability of occurrence. Four fuel streams are identified – natural gas, flare gas-A/B, flare gas-C and pure hydrogen. Ranked nodes are used for "Fuel Gas", "Hydrogen Carryover" and "Flare Frequency". As flare-C and natural gas does not have hydrogen in their composition, they are categorized in none or low hydrogen, whereas flare-A and flare-B has nearly 42 mole% hydrogen from collected data, so they are categorized into medium to low hydrogen composition. Pure hydrogen is categorized as high hydrogen carryover. Weightage factor of wmax (6.0 Fuel Gas,0.05 Flare Frequency) is used for parent node “H2 Carryover” node, showing the effect of hydrogen in firebox more on flare composition, even if frequency is relatively less. The advantage of Bayesian Network is, in an abnormal flaring, if the flare gas hydrogen composition and frequency vary for individual streams, then the child nodes- “Fuel Gas” and “Flare Frequency” node can be updated with different expressions and categorization (or states) to get an updated likelihood of consequence.

4.4.1. Sensitivity Analysis in Bayesian Network

In Figure 36, natural gas is chosen, to evaluate the likelihood of top events and their consequences in a year. Similarly, flare-A, flare-B and base case natural gas were chosen to check the likelihood of occurrence in a year. Figure 37 and Figure 38 shows the top event and consequence probability in a year for flare-C as fuel gas.

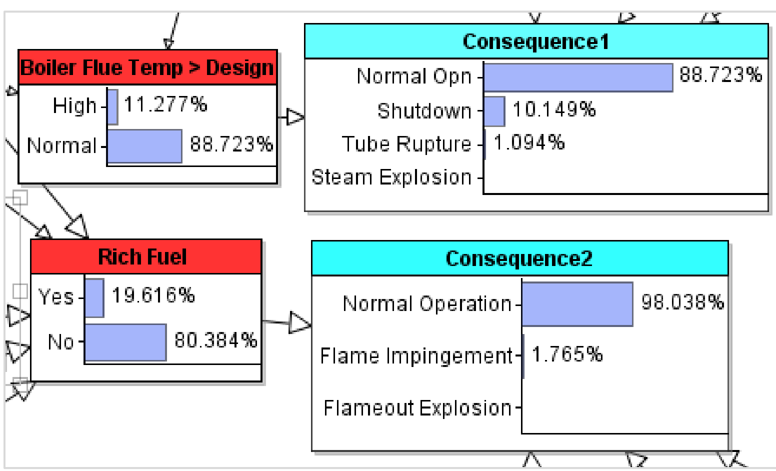


Figure 37 Frequencies of Top Events and Consequences for Base Case Boiler Operation

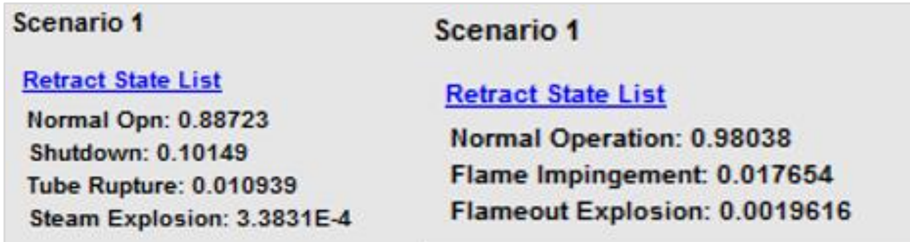


Figure 38 Probability of Occurrence of Critical Events for Base as Fuel

After sensitivity analysis, the probability occurrence of top events and consequences (in a year) for each stream are tabulated as shown in Table 26.

Table 26 Sensitivity Analysis of Top Events and Consequence Nodes

Conditions	Natural Gas/ Flare-C	Flare-A/B	Pure Hydrogen
Boiler Temp> Design (TE)	0.113	0.123	0.311
Normal Operation	0.887	0.877	0.689
Shutdown	0.101	0.111	0.280
Tube Rupture	1.09E-02	1.19E-02	3.02E-02
Steam Explosion	3.38E-04	3.68E-04	9.33E-04
Rich Fuel in Firebox (TE)	0.196	0.223	0.728
Normal Operation	0.980	0.978	0.927
Flame Impingement	1.765E-02	2.008E-02	6.554E-02
Flameout/ Stack Explosion	1.962E-03	2.231E-03	7.282E-03

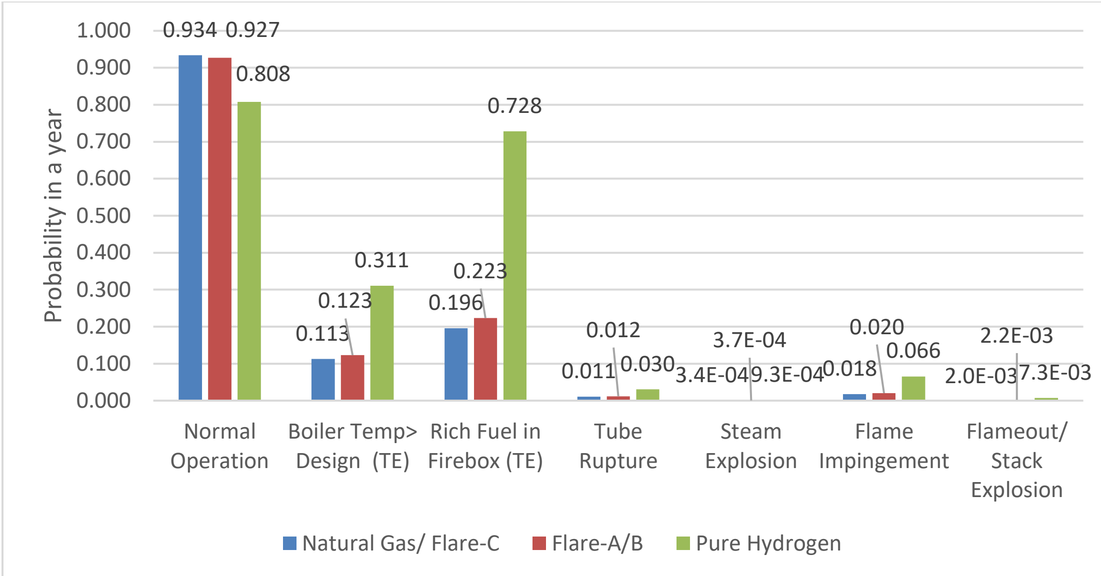


Figure 39 Change in Probability of Events for Flare Gas with Variable Composition

Figure 39 shows the significant effect on the boiler operational time, loss control events and its consequence on tube rupture and flame impingement. Considering these are average probability values, there is a significant rise in boiler gas temperature and rich fuel gas in firebox due to change to higher hydrogen fuel.

In a year, the operational time of steam boiler was observed to be 0.93, which has reduced, by 0.007 with flare-A/B as fuel and 0.126 with pure hydrogen as fuel. The likelihood of top event, boiler gas exceeding design temperature is 0.11 for base case, which has increased by 0.010 for flare-A/B and 0.20 for pure hydrogen. In addition, the chances of rich fuel in firebox was 0.196, which has increased by 0.027 for flare-A/B and by a huge difference of 0.53 for pure hydrogen. Noticeable consequences from top events are rupture of boiler syphon tubes and flame impingement. In base case, the probability of occurrence of tube rupture and flame impingement was 0.011 and 0.018. However, in case of using pure hydrogen to the boiler fuel, the respective probabilities have increased by 0.019 and 0.048. For, other consequences, the probability values are marginal. However, if the flare gas frequency increases from 0.033 (12 hrs/year) during abnormal flaring, the consequences will change.

Thus, following inferences can be made from Bayesian Network:

- The change of fuel from base case to flare gas –A/B, reduces the probability of operational time by 0.007, while increasing the boiler gas temperature by 0.010 and rich fuel condition in boiler by 0.027.
- There is a significant effect of pure hydrogen on safety constraints and their consequences. The operational time reduces by 0.126. There is a high probability of boiler temperature exceeding design limit (probability change is 0.20) and rich fuel in firebox (probability change is 0.532).
- Pure hydrogen as fuel, increases the probability of tube rupture by 0.02 and flame impingement by 0.048.
- The occurrence of steam explosion, flameout/stack explosion are very low (below 7.5E-03). If the operating life of fire tube boiler is considered 30 years, then the probability of steam explosion and flameout/stack explosion are less than 0.25.
- The probability of occurrence of incidents are marginal for flare-A and flare-B. However, if the frequency of flare gas utilization is increased, the consequences will change.

5. CONCLUSIONS AND FUTURE WORK

5.1. Conclusions

In this study, the effect of flare gas quality on cogeneration system is investigated after integration of flare gas to cogeneration fuel system. The objective of the study was to identify the hazards on integrating flare streams to an existing boiler, evaluate the changes in operating conditions and to determine the possibility of risk escalation due to changes in fuel quality.

The study developed a framework to identify process hazards and process upsets through process simulation and sensitivity analysis, followed by scenario development in fault tree and event tree, which were subsequently mapped to Bayesian Network to evaluate cause and effect relationship and determine the risk escalation due to changes in boiler fuel quality. The uniqueness of the proposed methodology is the identification of process hazards and their associated operational effects through process simulation of base case and with the change in operating parameters, i.e. variable fuel quality and temperature.

From process simulation, flare stream temperature and presence of higher molecular weight hydrocarbons in flare streams showed minimal effect on boiler operation with constant fuel mass flow and pressure. Natural gas in base condition starts forming condensate at temperatures below -35°F, which leads to loss of energy. However, high hydrogen content has significant effect on boiler operating conditions. Increase in the hydrogen content in flare affects the boiler gas temperature and combustion mixture in the firebox. Higher concentrations of hydrogen above 60 mole % with ethylene and 75% of hydrogen with methane, elevated radiation zone duty and gas temperature is predicted. Moreover, at constant air mass flow, lighter hydrocarbons in fuel system leads to incomplete combustion of fuel gas in firebox and results in carryover of unburnt hydrocarbons to flare stack, which have the potential to ignite and explode at elevated temperature. Pure hydrogen when at stoichiometric air to fuel ratio at the given base case condition, increases the boiler radiation zone duty and process steam temperature by a huge margin, rendering mechanical damage to tubes and burner nozzles.

On development of incident scenarios and quantitatively assessing the technical risk in Bayesian Network, the escalation of risk due to hydrogen carryover in fuel were significant. The occurrence of boiler gas temperature exceeding design limit and rich fuel in boiler firebox has increased by a margin of more than 0.10 and 0.027 per year for flare streams containing average hydrogen, while for higher hydrogen content fuel, the respective probabilities have increased by 0.20 and 0.53. Though, the change in likelihood of flame impingement on tubes, tube rupture and flameout or stack explosion incidents are marginal for medium hydrogen content flare gas (below 0.002), the change in likelihood of flame impingement and tube rupture are higher (0.048 and 0.02) for high hydrogen content gas. An important conclusion that can be drawn from this work is that the presence of pure hydrogen in flare gas has significant impact on boiler operation and can lead to loss control events and untoward incidents. To utilize higher hydrogen content flare gas for cogeneration, process needs to update preventive safeguards to avert the occurrence of top events and increase the operational time of steam boiler.

5.2. Future Work

For future work, the steam boiler base model to be run on dynamic simulation to investigate the effect of flare gas quality on steam boiler fuel controller, when the fuel control valve is in auto control with process steam outlet temperature. Secondly, in dynamic simulation, the effect of high combustion in firebox on fired boiler pressure profile (high draft can lead to gas loss through arch section). Finally, the consequence analysis of stack explosion and steam explosion to understand the severity of the incidents identified and based on the risk acceptance, the adequacy of preventive safeguards (e.g. Wobbe Calorimeter, NOx burners).

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